Retrofitting Municipal Wastewater Treatment Plants for Enhanced Biological Phosphorus Removal

Report EPS 3/UP/3
October 1986

Environment CANADA Environnement
20410601 VOL ISS 3 861000
EXC # 001
TD 169 R463
REPORT (CANADA. ENVIRONMENTAL PROTE
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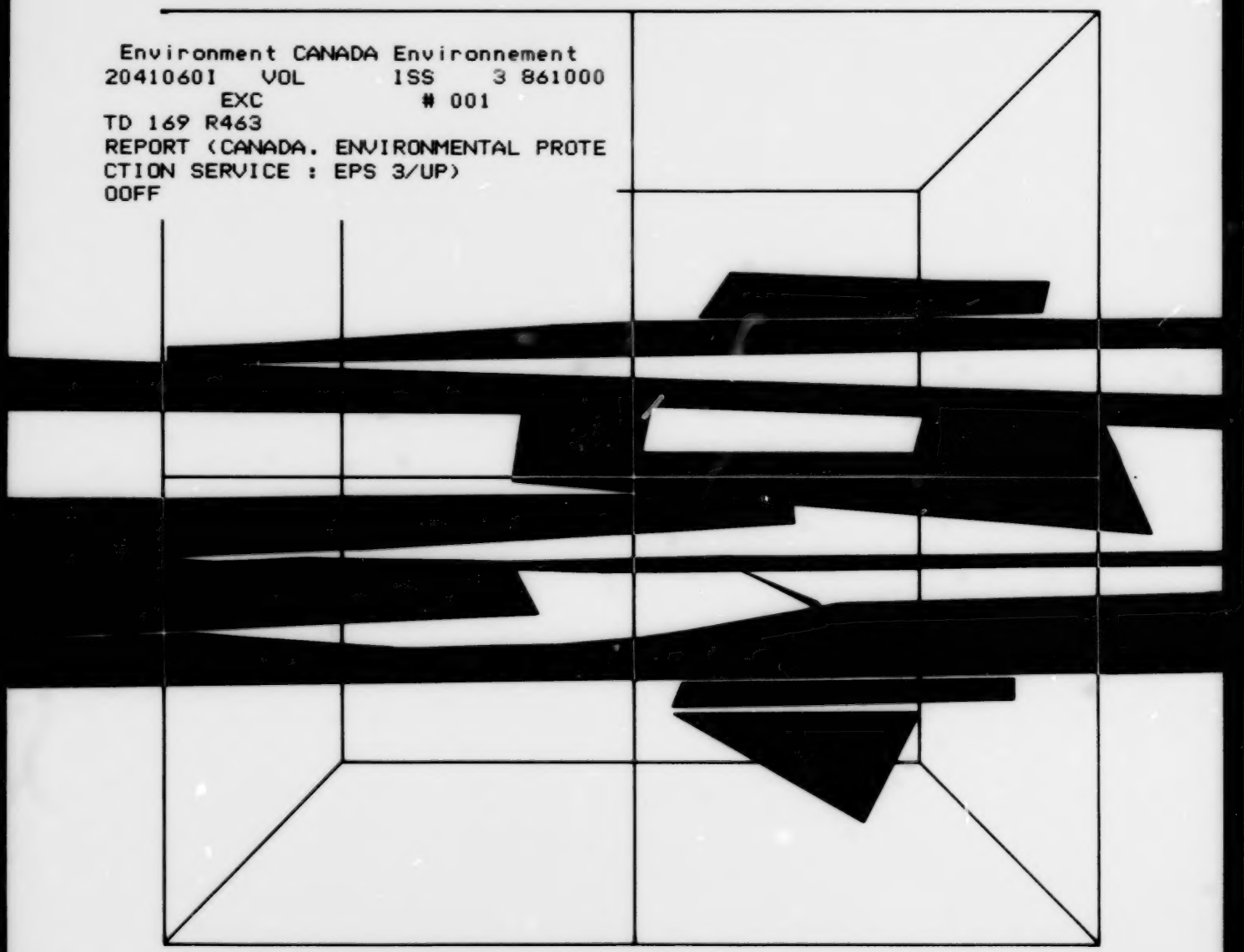

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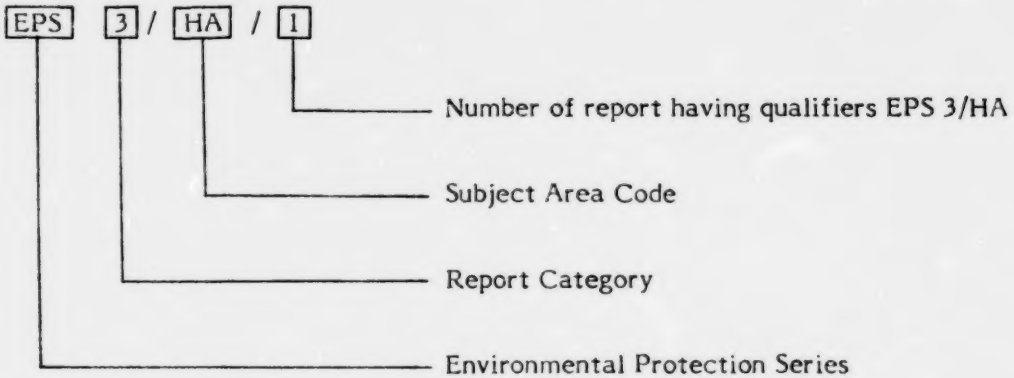

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RETROFITTING MUNICIPAL WASTEWATER TREATMENT PLANTS FOR ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL

by

CANVIRO Consultants Ltd.
Kitchener, Ontario

with

Norbert W. Schmidtke and Assoc. Ltd.
Kitchener, Ontario

and

David I. Jenkins and Assoc. Inc.
Berkeley, California

for the

Environmental Protection Programs Directorate
Conservation and Protection
Environment Canada

Report EPS 3/UP/3
October 1986

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Dr. Henryk Melcer
Wastewater Technology Centre
Environment Canada
P.O. Box 5050
Burlington, Ontario
L7R 4A6

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Section des publications
Direction générale des programmes de protection de l'environnement
Environnement Canada
Ottawa, Ontario
K1A 0E7

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Cat. No. En 49-8/3-3E
ISBN 0-662-15051-1
BEAUREGARD PRESS LIMITED

ABSTRACT

The technical and economic feasibility of modifying existing municipal wastewater treatment plants in Canada for enhanced biological phosphorus removal (EBPR) was evaluated. The cost of building new plants incorporating EBPR technology was also assessed. Cost analyses were based on three processes: primary treatment, conventional activated sludge (CAS) and extended aeration, with designs based on standard practice. Effluent total phosphorus limits of 1.0 and 0.3 mg/L were compared. Four EBPR processes, Bardenpho, UCT (University of Capetown), Air Products' A/O Process, and Phostrip, were addressed. The technical review concluded that chemical precipitation and/or effluent filtration were required with the EBPR processes to achieve consistent effluent phosphorus limits of 1.0 mg/L or less. Further, it was found that, ideally, EBPR and chemical precipitation processes should be applied simultaneously for optimum performance and minimum cost.

The economic review concluded that, of the model plants, CAS systems were the most suitable candidates for modifications to EBPR technology. EBPR processes typically required higher capital investment than chemical precipitation. However, the size at which the EBPR processes became cost-competitive with chemical precipitation was a function of the process complexity. In terms of annual operation and maintenance (O&M) costs, all EBPR options except Phostrip were cost competitive with chemical precipitation at flows greater than 36.4 ML/d. Based on total annual O&M costs, only the A/O process was more cost-effective in every retrofit situation evaluated because of the additional cost of amortizing the capital expenditure.

In new facilities, CAS plants incorporating chemical precipitation were more cost-effective than all the EBPR processes except A/O, based on total annual costs. However, on the basis of annual O&M costs, A/O was generally more cost-effective than the conventional plant at the 1.0 mg/L effluent phosphorus level. Furthermore, all EBPR processes except Phostrip had lower annual O&M costs than the conventional plant at the 0.3 mg/L effluent phosphorus level and at flows greater than 13.6 ML/d.

The assessment of technologies for phosphorus removal should not be based on economic evaluations alone. Other important factors that should also be addressed in retrofit situations include the desired effluent nutrient limits, the design of existing facilities and the characteristics of wastewaters in Canada. Circumstances that would make EBPR processes more attractive in Canada include more stringent phosphorus and nitrogen removal requirements, major increases in chemical costs and greater confidence in the performance of EBPR processes.

RÉSUMÉ

Le présent rapport évalue la faisabilité technique et économique de la conversion des stations canadiennes d'épuration des eaux usées urbaines à l'élimination biologique améliorée du phosphore (EBAP), ainsi que les coûts de construction de nouvelles stations intégrant les procédés d'EBAP. Les analyses de coûts ont porté sur trois procédés: le traitement primaire, les boues activées classiques (BAC) et l'aération prolongée, auxquels on a attribué des caractéristiques de conception standards. Pour fins de comparaison des coûts, des teneurs limites en phosphore total du rejet de 0,3 et 1,0 mg/L ont été utilisées. Quatre procédés d'EBAP, soit le Bardenpho, l'UCT (University of Capetown), l'A/O d'Air Products et le Phostrip, ont été examinés. Une analyse technique a mené à la conclusion que les procédés d'EBAP permettaient d'atteindre une teneur limite en phosphore du rejet de 1,0 mg/L ou moins, à la condition d'être appuyés par une précipitation chimique et/ou par une filtration de leur effluent. Il a également été constaté que les procédés d'EBAP et de précipitation chimique devraient idéalement agir simultanément de façon à optimiser le rendement et à réduire les coûts au minimum.

Une analyse économique a indiqué que, de toutes les stations modèles, celles qui utilisaient des systèmes à BAC convenaient le mieux aux modifications requises par la technique de l'EBAP. Autre constatation, les procédés d'EBAP exigent d'ordinaire des immobilisations plus considérables que la précipitation chimique. Il faut dire, cependant, que la taille de la station à partir de laquelle les coûts d'un procédé d'EBAP concurrencent ceux de la précipitation chimique sont fonction de la complexité du procédé. En ce qui concerne les frais annuels d'exploitation et d'entretien, toutes les variantes d'EBAP, à l'exception du procédé Phostrip, font aussi bonne figure que la précipitation chimique lorsque le débit hydraulique de la station dépasse 36,4 millions de L/j. En tenant compte des frais annuels globaux d'exploitation et d'entretien, seul le procédé A/O s'est révélé plus rentable que la précipitation chimique dans tous les cas de conversion étudiés, à cause du coût additionnel moindre de l'amortissement des immobilisations.

Dans les stations nouvelles, les traitements classiques par boues activées comportant une précipitation chimique se sont révélés plus rentables que les procédés d'EBAP (exception faite du procédé A/O) si l'on prend comme critère les frais annuels globaux. Cependant, si l'on considère les seuls frais annuels d'exploitation et d'entretien, le procédé A/O est généralement plus rentable que le traitement classique lorsqu'il s'agit d'atteindre une teneur limite en phosphore de 1,0 mg/L. En outre de cela, les procédés d'EBAP (exception faite du Phostrip) entraînent des frais annuels d'exploitation et

d'entretien inférieurs à ceux du traitement classique lorsque la teneur limite en phosphore à atteindre est de 0,3 mg/L et que le débit hydraulique de la station dépasse 13,6 millions de L/j.

L'évaluation des techniques de déphosphoration ne devrait pas être restreinte à la dimension économique. D'autres éléments devraient être pris en considération lorsque l'on envisage la possibilité d'une conversion, notamment la teneur en nutriants du rejet requise, la conception des installations en place et les caractéristiques des eaux usées au Canada. Quant aux facteurs susceptibles de rendre les procédés d'EBAP plus intéressants dans le contexte canadien, mentionnons l'adoption de normes de rejet plus strictes concernant le phosphore et l'azote, des hausses substantielles du prix des agents chimiques et bien sûr une augmentation de la confiance en l'efficacité de ces procédés.

TABLE OF CONTENTS

		Page
ABSTRACT		iii
LIST OF FIGURES		ix
LIST OF TABLES		xi
1	INTRODUCTION	1
1.1	Study Objectives	1
1.2	Scope of Work	2
1.3	Report Organization	3
2	TECHNICAL ASSESSMENT OF PHOSPHORUS REMOVAL TECHNOLOGIES	4
2.1	Enhanced Biological Phosphorus Removal	4
2.1.1	Process Alternatives	5
2.1.2	Process Descriptions	9
2.1.3	Process Comparisons	10
2.1.4	Design Criteria	12
2.1.5	Process Performance Data	16
2.1.6	Sludge Generation Data	23
2.1.7	Process Sensitivity and Response to Low Temperature	23
2.1.8	Process Response to Organic, Toxic and Hydraulic Shock and Upset Conditions	24
2.1.9	Operating Problems	25
2.1.10	Technical Feasibility of EBPR Processes	27
2.1.11	Summary	30
2.2	Chemical Phosphorus Removal	31
2.2.1	Process Alternatives	31
2.2.2	Process Descriptions	31
2.2.3	Process Design Criteria	32
2.2.4	Process Performance Data	33
2.2.5	Sludge Generation Data	33
2.2.6	Process Sensitivity, Response and Recovery from Upset Conditions	33
2.2.7	Operating Problems	34
2.2.8	Technical Feasibility	34
3	STUDY METHODOLOGY	35
3.1	Design of Model Plants	35
3.1.1	Model Primary Plant	35
3.1.2	Model Conventional Activated Sludge Plant	35
3.1.3	Model Extended Aeration Plant	35
3.1.4	Size of Model Plants	38
3.2	Costing of Model Plants	40
3.2.1	CAPDET Simulation	40
3.2.2	Cost Bases for Model Plants	41
3.3	Retrofits to Achieve Phosphorus Control	42

		Page
3.3.1	Design of Chemical Phosphorus Removal Facilities	42
3.3.2	Design of EBPR Processes	42
3.3.3	Retrofit Process Equipment List	44
3.3.4	Development of Retrofit Costs	44
3.3.5	Cost Bases for Retrofit Plants	45
3.4	New Plant Costs	58
4	ECONOMIC ASSESSMENT OF PHOSPHORUS REMOVAL TECHNOLOGIES	59
4.1	Retrofit Costs	59
4.1.1	Chemical Precipitation Process	59
4.1.2	Bardenpho Process	60
4.1.3	UCT Process	63
4.1.4	A/O Process	66
4.1.5	Phostrip Process	71
4.1.6	Summary of Retrofit Costs	74
4.2	New Plant Costs	74
4.2.1	Chemical Precipitation	77
4.2.2	Bardenpho Process	77
4.2.3	UCT Process	77
4.2.4	A/O Process	78
4.2.5	Phostrip Process	78
4.2.6	Summary of New Plant Costs	81
4.3	Cost Sensitivity Analysis	83
4.3.1	Effect of Chemical Dosage and Chemical Cost	84
4.3.2	Effect of Chemical Type	85
4.4	Accuracy of Cost Estimates	85
5	POTENTIAL APPLICATIONS OF EBPR PROCESS RETROFITS IN CANADA	88
5.1	Effluent Phosphorus Limits	88
5.2	Effluent Nitrogen Limits	88
5.3	Wastewater Characteristics	89
5.4	Characteristics of Existing Canadian WPCPs	90
5.5	Sludge Management Implications	91
5.6	State of Development of Phosphorus Removal Technologies	93
5.7	Avenues of Capital Funding	94
5.8	Summary	94
6	CONCLUSIONS	96
6.1	Technical Feasibility of EBPR Processes	96
6.2	Economics of EBPR Process Retrofits	96
6.3	Economics of New EBPR Facilities	97
6.4	Applicability of EBPR Process Retrofits in Canada	97
REFERENCES		99
APPENDIX	COST FUNCTIONS	105

LIST OF FIGURES

Figure		Page
1	ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL PROCESS ALTERNATIVES	7
2	PHOSTRIP PROCESS ALTERNATIVES	8
3	MODEL PRIMARY PLANT PROCESS FLOWSHEET	36
4	MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT PROCESS FLOWSHEET	37
5	MODEL EXTENDED AERATED PLANT PROCESS FLOWSHEET	39
6	RETROFIT OF MODEL PRIMARY PLANT TO BARDENPHO PROCESS	46
7	RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO BARDENPHO PROCESS	47
8	RETROFIT OF MODEL EXTENDED AERATION PLANT TO BARDENPHO PROCESS	48
9	RETROFIT OF MODEL PRIMARY PLANT TO UCT PROCESS	49
10	RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE TO UCT PROCESS	50
11	RETROFIT OF MODEL EXTENDED AERATION PLANT TO UCT PROCESS	51
12	RETROFIT OF MODEL PRIMARY PLANT TO A/O PROCESS	52
13	RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO A/O PROCESS	53
14	RETROFIT OF MODEL EXTENDED AERATION PLANT TO A/O PROCESS	54
15	RETROFIT OF MODEL PRIMARY PLANT TO PHOSTRIP PROCESS	55
16	RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO PHOSTRIP PROCESS	56
17	RETROFIT OF MODEL EXTENDED AERATION PLANT TO PHOSTRIP PROCESS	57
18	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR BARDENPHO AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS	64

Figure

		Page
19	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR BARDENPHO AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS	64
20	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR UCT AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS	67
21	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR UCT AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS	67
22	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR A/O AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS	70
23	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR A/O AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS	70
24	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR PHOSTRIP AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS	73
25	COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR PHOSTRIP AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFFFLUENT TOTAL PHOSPHORUS	73
26	COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)	81
27	COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)	82
28	COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT	82

LIST OF TABLES

Table		Page
1	FIVE-STAGE BARDENPHO BASIN HYDRAULIC DETENTION TIMES	13
2	TYPICAL A/O AND A^2/O DESIGN AND OPERATING PARAMETERS	14
3	SUMMARY OF PHOSTRIP PERFORMANCE (NORTH AMERICA)	17
4	SUMMARY OF BARDENPHO PERFORMANCE (NORTH AMERICA)	18
5	SUMMARY OF A/O PERFORMANCE (NORTH AMERICA)	18
6	PROCESS OPTIONS FOR CURRENTLY OPERATING EBPR PLANTS IN NORTH AMERICA	19
7	ESTIMATED EFFLUENT TOTAL PHOSPHORUS FROM EBPR PROCESSES	20
8	30-DAY AVERAGE AND DAILY MAXIMUM TOTAL P CONCENTRATIONS AT THE TAHOE TRUCKEE SANITATION AGENCY, CA, PHOSTRIP PLANT	21
9	30-DAY AVERAGE AND DAILY MAXIMUM EFFLUENT TOTAL P AND SS CONCENTRATIONS AT THE ROCHESTER, MN, PHOSTRIP PLANT	22
10	SUMMARY OF PONTIAC, MI, A/O PROCESS PERFORMANCE	23
11	A/O PILOT PLANT PERFORMANCE AT PATAPSCO UNDER INFLUENT FLOW VARIATION	24
12	EBPR PROCESS REQUIREMENTS FOR VARIOUS EFFLUENT TOTAL PHOSPHORUS LIMITS	28
13	EFFECT OF CHEMICAL PHOSPHORUS REMOVAL ON SLUDGE GENERATION RATES	33
14	DISTRIBUTION OF TREATMENT PLANTS IN ONTARIO ACCORDING TO TYPE AND SIZE	38
15	INDEX OF EBPR RETROFIT PROCESS DESIGNS	44
16	COSTS TO RETROFIT PRIMARY PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL	60
17	COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL	61
18	COSTS TO RETROFIT EXTENDED AERATION PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL	61

Table		Page
19	COSTS TO RETROFIT PRIMARY PLANTS WITH BARDENPHO TECHNOLOGY	62
20	COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH BARDENPHO TECHNOLOGY	62
21	COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH BARDENPHO TECHNOLOGY	62
22	COSTS TO RETROFIT PRIMARY PLANTS WITH UCT TECHNOLOGY	65
23	COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH UCT TECHNOLOGY	65
24	COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH UCT TECHNOLOGY	65
25	COSTS TO RETROFIT PRIMARY PLANTS WITH A/O TECHNOLOGY	68
26	COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH A/O TECHNOLOGY	68
27	COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH A/O TECHNOLOGY	68
28	COSTS TO RETROFIT PRIMARY PLANTS WITH PHOSTRIP TECHNOLOGY	72
29	COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH PHOSTRIP TECHNOLOGY	72
30	COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH PHOSTRIP TECHNOLOGY	72
31	SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)	75
32	SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)	75
33	SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 0.3 mg/L TOTAL PHOSPHORUS LIMIT	75
34	SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)	76

Table		Page
35	SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)	76
36	SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT	76
37	NEW FACILITY COSTS - CONVENTIONAL ACTIVATED SLUDGE PLUS CHEMICAL ADDITION	77
38	NEW FACILITY COSTS - BARDENPHO TECHNOLOGY	79
39	NEW FACILITY COSTS - UCT TECHNOLOGY	79
40	NEW FACILITY COSTS - A/O TECHNOLOGY	80
41	NEW FACILITY COSTS - PHOSTRIP TECHNOLOGY	80
42	SUMMARY OF TOTAL ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)	83
43	SUMMARY OF ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)	84
44	SUMMARY OF ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT	84
45	SENSITIVITY OF TOTAL ANNUAL COSTS TO CHEMICAL DOSAGE AND CHEMICAL COST	85
46	SENSITIVITY OF ANNUAL O&M COSTS TO CHEMICAL DOSAGE AND CHEMICAL COST	86
47	SENSITIVITY OF TOTAL ANNUAL COSTS TO CHEMICAL TYPE	86
48	CHARACTERISTICS OF RAW WASTEWATERS IN ONTARIO	91

1 INTRODUCTION

Phosphorus removal has been practised at most municipal wastewater treatment plants discharging to the Lower Great Lakes Basin since the late 1970s. In other areas where receiving waters are sensitive to nutrient load, phosphorus removal requirements also have been implemented. Control of ammonia and/or oxidized nitrogen compounds (nitrate and nitrite) has been site-specific.

Traditionally, phosphorus removal strategies have incorporated chemical precipitation by the addition of metal salts (alum, ferric chloride, lime). Although this technology is well-proven for achieving effluent discharge levels of 1 mg/L total phosphorus (TP) or less, costs increase significantly as more stringent requirements are placed on effluent total phosphorus concentrations. Furthermore, chemical precipitation alone does not provide control for nitrogenous compounds in effluents.

Several new wastewater treatment processes have been developed to biologically achieve a high degree of phosphorus removal. In some instances, nutrient discharges (nitrogen and phosphorus) can be completely controlled using these configurations. These processes, although relatively unproven in Canada, appear to offer advantages relative to conventional chemical precipitation technology. However, the technologies are complex compared to the conventional practice. As well, the economic benefits, particularly in retrofit situations, can be very site-specific.

This investigation addressed the technical and economic feasibility of retrofitting existing Canadian municipal wastewater treatment plants for enhanced biological phosphorus removal (EBPR). Various EBPR processes were reviewed briefly and the practicality of using these technologies compared to conventional chemical precipitation technology for removing phosphorus from municipal wastewaters was assessed. An attempt was made to identify the type and size of existing treatment plants in Canada that would be most amenable to retrofit for enhanced biological phosphorus removal. Costs were compared for a number of EBPR alternatives and chemical precipitation. This study was not limited to a single specific plant, but was conducted in a broader context so that those conditions under which either chemical or enhanced biological phosphorus removal was more cost-effective could be identified.

1.1 Study Objectives

The overall objective of the study was to examine the technical and economic feasibility of retrofitting existing Canadian municipal wastewater treatment plants for enhanced biological phosphorus removal (EBPR).

Specific objectives of the study were:

- to compare the capital and O&M costs associated with converting existing treatment facilities to EBPR processes with the costs of implementing chemical precipitation technologies at these facilities;
- to develop similar cost comparisons between biological and chemical phosphorus removal process alternatives for new treatment facilities;
- to develop an approach, including an accessories list with cost factors for the various components, that could be utilized by design engineers to estimate the retrofit costs associated with existing wastewater treatment plants;
- to comment on the applicability of EBPR processes in Canada.

1.2 Scope of Work

In order to streamline the investigation and reduce the number of retrofit situations requiring cost analysis to a manageable level, the scope of work was limited to the following situations:

i) Three types of plants were to be considered: primary plants, conventional activated sludge plants and extended aeration plants. High-rate activated sludge plants were eliminated because of the limited use of this technology in Canada.

ii) Each plant type was evaluated at three levels of flow. The flow rates selected were the lower quartile, median and upper quartile flows based on frequency distributions of flow rates for each type of plant in Ontario.

iii) The biological nutrient control technologies to be evaluated were to include the three commercially available processes used in North America - Bardenpho, Phostrip, and A/O - as well as the developmental University of Cape Town (UCT) process.

iv) Ferric chloride was used to develop comparative costs for chemical precipitation technology due to its widespread use in Ontario. A sensitivity analysis was conducted to determine the impact of chemical cost and chemical type on the cost data.

v) Two total phosphorus (TP) discharge limits were imposed on the effluents from the treatment facilities: 1.0 mg/L and 0.3 mg/L TP. However, as it was not considered technically feasible to achieve the lower limit (0.3 mg/L) by chemical addition to a

primary plant due to the high suspended solids concentrations in primary effluents, the cost analysis was conducted only to a 1.0 mg/L TP limit for the primary plants. Cost analyses of alternatives to achieve TP discharge limits of 0.5 mg/L were eliminated because design criteria and process knowledge of the biological options are not sufficiently developed to discriminate between design effluent limits of 0.3 mg/L and 0.5 mg/L.

vi) The raw wastewater characteristics were assumed to be:

COD	=	250 mg/L
TKN	=	25 mg/L
Alkalinity	=	240 mg/L
Total Phosphorus	=	5 mg/L

1.3 Report Organization

Section 2 of the report presents the results of the technical review of phosphorus removal technologies, both biological and chemical. The findings of this review provided the basis for the designs used for the development of capital and operating cost estimates for phosphorus removal technologies.

Section 3 summarizes the methodology used in the design and costing of phosphorus control options.

Section 4 presents the result of the cost analysis for phosphorus removal process alternatives, for both retrofit plants and new plants.

Section 5 reviews existing nutrient removal requirements in Canada and presents a discussion of the applicability of EBPR technologies within the context of the technical and economic data previously presented. The conclusions resulting from the investigation are presented in Section 6.

Cost curves for the accessories required to retrofit existing municipal wastewater treatment plants for phosphorus removal are presented in the Appendix. The cost curves are the specific curves used for the plant retrofits described in Section 4 of the report, and those for additional process equipment (e.g., dissolved air flotation thickeners, flow equalization facilities) that might be required for a specific installation.

2 TECHNICAL ASSESSMENT OF PHOSPHORUS REMOVAL TECHNOLOGIES

2.1 Enhanced Biological Phosphorus Removal

The nutrient requirements (nitrogen and phosphorus) for biological treatment of wastewaters classically are stated in terms of the BOD_5/N and BOD_5/P ratios. For satisfactory treatment, a $BOD_5/N/P$ ratio of approximately 100:5:1 is required. For the typical yields of sludge produced by biological treatment processes, these nutrient requirements translate into sludge phosphorus contents in the range from 1.5% to 2.5% P/volatile suspended solids (dry weight basis). In the context of typical domestic sewage composition and treatment processes (e.g., primary sedimentation, biological treatment, sludge digestion with supernatant return), phosphorus is reduced by approximately 2 mg P/L, or 10% to 30% depending on influent phosphorus concentrations. These are the expected norms for phosphorus removal by biological waste treatment processes. Some biological waste treatment processes, however, achieve much greater phosphorus removal without the addition of phosphate precipitating chemicals. This phenomenon, which has become embodied in several process flowsheets, is referred to here as enhanced biological phosphorus removal (EBPR). While the mechanisms of EBPR are not fully understood, the following process conditions and observations present a general picture that is supported by the available data and is consistent with the principles of microbial competition in activated sludge.

a) Return activated sludge (RAS) and influent waste must mix initially and be allowed to react under anaerobic conditions. "Anaerobic" is defined in waste treatment as the condition in which no nitrate (NO_3^-) or dissolved oxygen (DO) is present.

b) In the initial anaerobic basin, sufficient time must be provided for the uptake of much of the soluble organic matter (soluble COD).

c) The anaerobic basin must be followed (at some point) by an aerated (oxic) basin.

d) In nitrifying activated sludge systems, provisions must be made to prevent return of NO_3^- to the anaerobic basin. To achieve this, denitrification is required.

e) In the anaerobic basin, orthophosphate is released from the sludge into solution and soluble COD is removed from solution.

f) In the aerobic basin, soluble orthophosphate is removed from solution; soluble COD does not increase.

g) Activated sludge exhibiting EBPR contains Neisser-positive staining granules ("volutin", polyphosphate). The granules decrease in number/size in the anaerobic basin compared to the aerobic basin.

h) Addition of readily utilizable soluble organic matter (e.g., acetate, septic sludge thickener overflow) to the influent waste aids the rate and extent of EBPR.

i) During the anaerobic phase, there is evidence that carbonaceous storage products (e.g., polyhydroxybutyrate) accumulate. During the subsequent oxic stages, they apparently are used by the microorganisms.

The picture that emerges from the above is that some members of the activated sludge community can gain a competitive advantage by virtue of their ability to take up soluble substrate under anaerobic conditions. This advantage is conferred by their ability to use the energy from breakdown of stored intracellular polyphosphate to orthophosphate to transport soluble organic matter into the cell and synthesize it into a carbonaceous storage product. Through this ability, these organisms preferentially acquire the soluble substrate and accumulate carbon reserves for use during the later stages of aerobic treatment when exogenous supplies of carbon are depleted. During the aerobic stages of treatment, these organisms restore their intracellular polyphosphate supply using soluble orthophosphate. Part of the stored carbonaceous substrate is used for this purpose.

When NO_3^- or DO are present at the point where RAS and influent wastewater mix, the polyphosphate-storing microorganisms do not compete successfully with microorganisms that use NO_3^- or DO as terminal electron acceptors. Thus, the maintenance of anaerobic conditions in the initial reaction tank is critical to the efficient operation of EBPR processes.

Many of the details of this global picture still need to be defined. For example, the nature of and pathways between storage products is not understood; the range of microorganisms types that participate in these reactions is not known; and the influence and importance of chemical species (e.g., metal ions) on EBPR is poorly understood. However, enough is known about the gross conditions associated with the success and failure of EBPR to allow the design (with the aid of preliminary pilot plant information) of EBPR processes.

2.1.1 Process Alternatives. Three commercial EBPR processes are appropriate for retrofitting to continuously fed activated sludge systems. These are Bardenpho, Phostrip and A/O (A^2/O). In North America, these processes are under license as follows:

Bardenpho	Eimco Process Equipment Co. 669 West Second South Salt Lake City. UT 84110
Phostrip	Biospherics Inc. 4928 Wyaconda Road Rockville, MD 20852
A/O, A^2/O	Air Products & Chemicals Inc. P.O. Box 538 Allentown, PA 18105

Figures 1 and 2 present idealized flowsheets for these processes. Labelling of the various process reactors has been made consistent to allow for comparison between processes.

Figures 1A to 1E illustrate the evolution of the Bardenpho process from its original development as a nitrogen-control process that achieved incidental phosphorus removal (1A). Subsequent developments were aimed at optimizing biological phosphorus uptake when it was recognized that anaerobic conditions were critical to the process (Figure 1B). The three-stage and two-stage processes (Figures 1C and 1D) allow for lesser degrees of nitrogen control in conjunction with biological phosphorus uptake. The UCT process (Figure 1E), developed at the University of Cape Town (Siebritz et al., 1982), was a modification of the Bardenpho process with additional internal recycle to ensure minimal nitrate concentrations in the anaerobic reactor.

The A/O and A^2/O processes are depicted in Figures 1F and 1G. The difference between the two process configurations relates to the degree of nitrogen control provided. A/O, a two-stage process, provides no nitrogen control and, in fact, nitrification in the A/O process will impair phosphorus removal as a result of nitrate recycle to the anaerobic reaction zone. A^2/O incorporates an anoxic reaction zone to achieve denitrification.

Figure 2 illustrates the available Phostrip process configurations. Unlike the Bardenpho and A/O processes, Phostrip is a sidestream process used in conjunction with a conventional suspended growth biological system, which may include nitrogen control depending on the specific requirements.

In addition to these EBPR processes, Bundgaard et al. (1983) has described an intermittently-fed process called Bio-Denipho, which achieves biological phosphorus removal. Sequencing batch reactor systems (SBRs) can also be designed with the proper sequence of anaerobic, anoxic and aerobic periods to achieve EBPR, and oxidation ditches have also been shown to be capable of providing biological phosphorus removal.

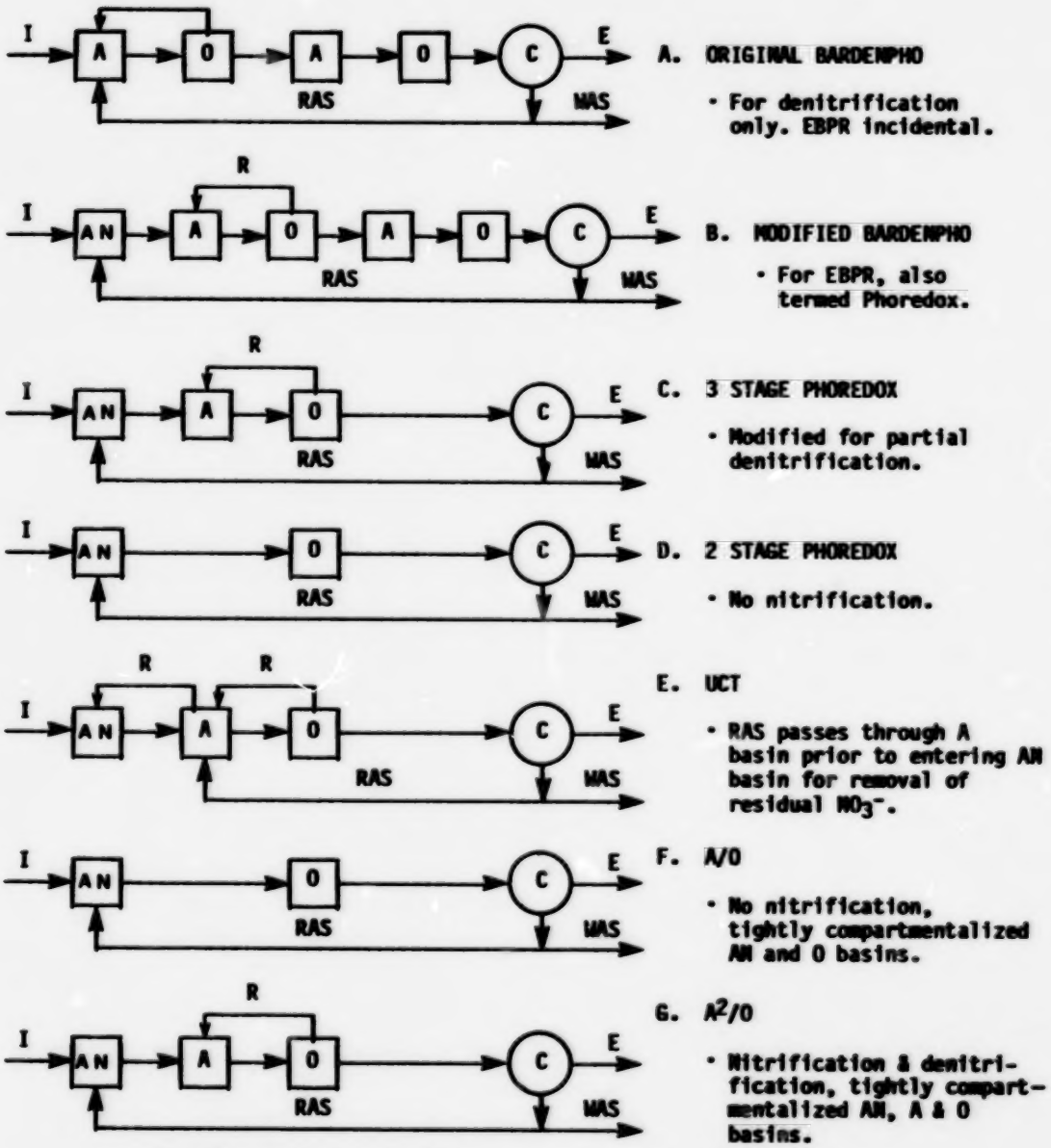
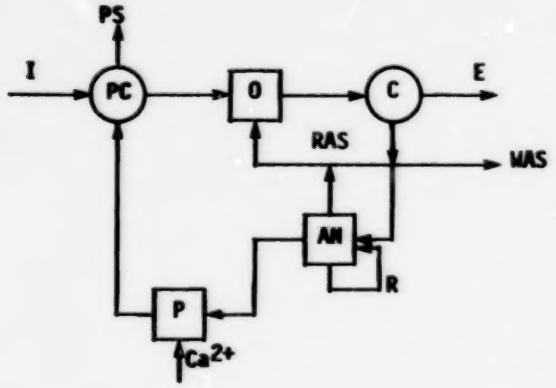
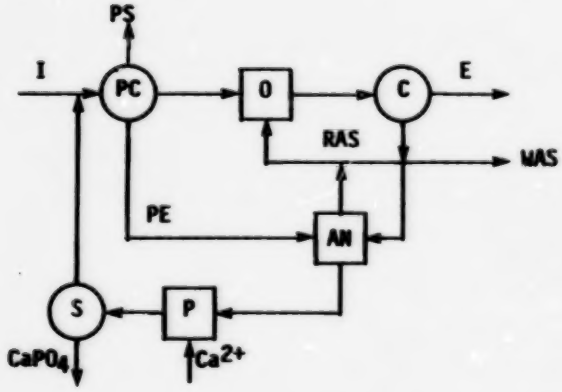


FIGURE 1 ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL PROCESS ALTERNATIVES



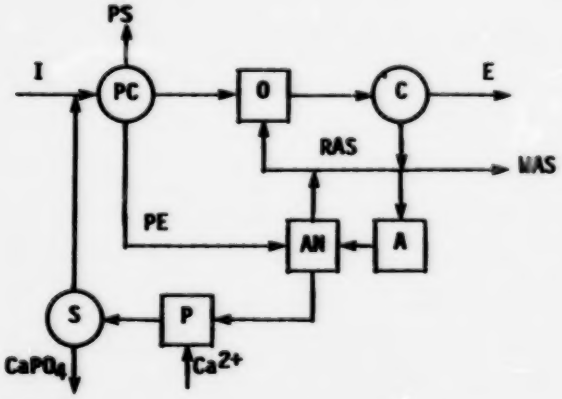
A. PHOSTRIP WITH RECYCLE STRIPPER (non-nitrifying system).

- Stripper (AN) overflow precipitated with lime & returned to PC for settling & removal with PC.



B. PHOSTRIP WITH ELUTRIATION STRIPPER (non-nitrifying system).

- Stripper (AN) "elutriated" with PE. Stripper overflow precipitated with lime (P) and settled. Chemical sludge disposed of separately. Overflow returned to PC.



C. PHOSTRIP FOR NITRIFYING ACTIVATED SLUDGE.

- Stripper (AN) is "protected" from NO_3^- in RAS by an anoxic denitrification basin (A).

RAS = Return Activated Sludge
WAS = Waste Activated Sludge
AN = Anaerobic Basin
A = Anoxic Basin
P = Reactor for Lime Addition
S = Lime Sludge Settler
R = Recycle Flow

I = Influent
E = Effluent
PC = Primary Clarifier
PS = Primary Sludge
PE = Primary Effluent
O = Aerobic Basin
C = Secondary Clarifier

FIGURE 2 PHOSTRIP PROCESS ALTERNATIVES

2.1.2 Process Descriptions. Process descriptions are provided for the three EBPR processes that are currently used in North America, and the developmental UCT process.

Bardenpho. This process is generally configured either as a three-stage or a five-stage system (Figure 1B and 1C). In the five-stage system, raw influent and RAS are fed to a mixed anaerobic basin. The mixture then flows to a mixed anoxic basin into which a recycle stream from the next downstream aerobic basin is introduced. From the anoxic basin, the mixed liquor flows in sequence to an aerated aerobic basin, a mixed anoxic basin, a final aerated aerobic basin and then to the secondary clarifier. The process objectives of the five-stage process are high nitrogen and phosphorus removal ($(NO_3^- + NO_2^- + NH_4^+)$-N < 1 mg/L; TP < 1 mg/L)). When less complete nitrogen removal can be tolerated ($(NO_3^- + NO_2^- + NH_4^+)$-N < 5 mg/L), a three-stage system is appropriate. The flowsheet for this process is identical to the five-stage system except that the last two stages (one anoxic and one aerobic) are deleted. These final two stages were originally incorporated to ensure complete nitrification and denitrification.

Bardenpho systems have been applied almost uniquely to nitrifying systems. However, there is no reason why they should not work in non-nitrifying plants, as indicated in Figure 1D for a two-stage system. In this configuration, any degree of nitrification will impair phosphorus removal due to the recycle of nitrate to the anaerobic reactor.

The various modifications that have been made to the original Bardenpho process to optimize phosphorus removal have resulted in considerable confusion in the nomenclature associated with the various flowsheets. The terms 'Bardenpho', 'modified Bardenpho', 'Phoredox', and 'modified Phoredox' have been applied with various meanings. For the purposes of this discussion, the processes shown in Figures 1A to 1D will be referred to generally as Bardenpho processes.

UCT. The UCT process (Figure 1E) can be considered to be a variation of the Bardenpho process. It differs only in that the RAS stream, rather than being fed directly to the initial anaerobic reactors, is routed through the anoxic reactor and, hence, to the anaerobic basin. This modification provides an additional barrier to the entry of NO_3^- into the anaerobic basin.

A/O and A^2/O. The flowsheet for the A^2/O process (Figure 1G) is identical to that for a three-stage Bardenpho process. This flowsheet would be applicable to a nitrifying system, where partial denitrification is required. In the A/O process (Figure 1F), influent waste and RAS flow first into a mixed anaerobic basin, then to an

oxygenated (or aerated) aerobic basin and finally to the secondary clarifier. This flowsheet is applicable to a non-nitrifying system and is identical to the two-stage Bardenpho process flowsheet mentioned previously. In detail, designs of Bardenpho and A/O systems have differed in the degree of compartmentalization of the basins (greater degree in A/O systems) and the use of high purity oxygen (A/O) rather than air (Bardenpho) for aeration.

Phostrip. The mainstream portion of the flowsheet (Figure 2) for the Phostrip process is the same as for a conventional activated sludge plant, i.e., primary clarifier, aeration basin and secondary clarifier. The RAS stream can be split so that some of it passes through an anaerobic basin called a phosphorus stripper. If the activated sludge process is nitrifying (Figure 2C), the RAS must first pass through a denitrification (anoxic, stirred) basin before entering the P-stripper, where phosphorus is elutriated from sludge. Two P-stripper flowsheets have been used. The recycle P-stripper removes released phosphate by recycling supernatant through the sludge blanket (Figure 2A). The elutriation P-stripper (Figure 2B) uses primary effluent to strip out released phosphate. Elutriation P-strippers are preferable because the primary effluent introduces soluble biodegradable organic matter which is required for anaerobic phosphate release. Additional elutriation liquid can be obtained by recycle of precipitation clarifier overflow through the P-stripper. Two streams leave the P-stripper. The underflow represents the phosphorus-deficient biological sludge which is recycled to the aeration basin. The overflow is the liquid containing the stripped (or elutriated) phosphate. The phosphate is removed from this stream by precipitation as calcium phosphate using lime. This process has been achieved in a variety of ways, including:

a) Lime addition to the stripper overflow in a mixed reactor followed by recycle of the entire reactor contents to the primary clarifier for co-settling and removal of calcium phosphate solids with the primary sludge (Figure 2A).

b) Lime addition to the stripper overflow followed by separation of the chemical sludge in a separate clarifier (Figures 2B and 2C); the supernatant is either recycled to the primary clarifier or used to aid in elutriation of the P-stripper. The chemical sludge is thickened separately prior to separate or combined dewatering and disposal with the primary and secondary waste sludges.

2.1.3 Process Comparisons. Some general comments with respect to the processes commercially used in North America are pertinent to provide an overview of EBPR technology.

a) The three-stage Bardenpho-based process flowsheet and A^2/O process flowsheet are basically identical.

b) The Bardenpho and A/O processes are "mainstream" processes in which all of the phosphorus removal processes occur in the same basins that treat the wastewater. By contrast, the Phostrip process (Figure 2) is a side-stream process in which the operations critical for EBPR are conducted on the RAS rather than in the liquid waste stream.

c) The Bardenpho and A/O processes do not indicate the need for any external chemical addition to achieve EBPR. In the Phostrip process, a chemical, usually lime, is added to precipitate the phosphorus released under anaerobic conditions.

d) In the Bardenpho and A/O processes, phosphorus can be removed only in the waste activated sludge. In the Phostrip process, phosphorus can be additionally removed in a chemical phosphate sludge stream.

e) The Bardenpho and Phostrip processes, being earlier applications of the principle of EBPR, show interesting process flowsheet developments that have taken place over time. This seems to be the result of an improved understanding of process mechanisms - especially the realization of the importance of initial anaerobic conditions and the presence of readily metabolizable soluble organic matter. The Bardenpho flowsheets, and ultimately the UCT flowsheet, demonstrate progressively greater precautions to prevent NO_3^- from entering the initial anaerobic basin. The Phostrip flowsheet changes from a recycle type stripper to an elutriation stripper in which the primary effluent is introduced into the stripper. This design change was made in response to the need for readily metabolizable soluble organic matter during the anaerobic phosphorus release phase. Further modifications of Bardenpho processes have been made (Oldham, 1984; Stensel et al., 1980) to achieve this end by such means as introducing septic sludge thickener overflow into the raw waste or allowing fermentation (acid digestion) of primary sludge in the primary clarifier.

f) Although not evident in the flowsheets in Figure 1 and not specifically addressed in any experimental work, it seems that the better performing EBPR plants of the Bardenpho and A/O type have highly compartmentalized basins (plug flow), especially in their anaerobic and anoxic sections. Such compartmentalization results in the exposure of the microorganisms to higher initial substrate concentrations and can also be used to confer additional flexibility of operation

(e.g., variation of the basin volumes devoted to anaerobic, anoxic and aerobic conditions). There also may be a possibility that compartmentalization in anaerobic and anoxic basins achieves better "isolation" of these processes, and also may indeed enhance the rates of the process.

2.1.4 Design Criteria.

Bardenpho. Three-stage and five-stage Bardenpho designs incorporate requirements for nitrification, denitrification and EBPR. Two design approaches are available, both of which have been developed using semi-empirical biokinetic models (Ekama et al., 1984; Barnard et al., 1984). Ekama et al.'s approach is:

a) Select incoming wastewater characteristics including soluble, readily biodegradable COD fraction, total kjeldahl nitrogen (TKN) and total phosphorus concentration.

b) Select range of practical sludge ages, minimum and maximum operating temperatures.

c) Calculate maximum unaerated (anaerobic plus anoxic) mass fraction, f_{xm}, to allow nitrification over entire range of expected operating temperatures.

d) Nitrification capacity can be estimated after an allowance for the nitrogen requirements of sludge synthesis.

e) The anaerobic mass fraction, f_{xa}, is selected, as well as an internal recirculation rate (typically 2:1 to 4:1).

f) The available denitrification capacity can be calculated from the size of the remaining anoxic mass fraction, f_{xdm}, where $f_{xdm} = f_{xm} - f_{xa}$ over the anticipated range of operating temperatures.

g) The remaining anoxic mass fraction, f_{xdm}, can be partitioned into a primary and secondary anoxic mass fraction based on an optimization procedure (this applies to the five-stage Bardenpho process).

h) The effluent nitrate can be calculated from the previously determined configuration of anoxic zones and recirculation rates.

i) The excess phosphorus removal propensity is determined by the soluble readily metabolizable COD concentration in the anaerobic zone and the hydraulic retention time in the same zone.

j) The effluent phosphate concentration is calculated from the difference between the influent total phosphorus concentration and the phosphorus requirement for the synthesis of new sludge plus the phosphorus stored in the cell.

k) The design procedure can be repeated for different values of the anaerobic mass fraction, f_{xa}, and the sludge age.

It should be noted that this design procedure predicts the effluent orthophosphate concentration and not the total phosphorus concentrations in the effluent.

Barnard et al.'s approach computes the minimum sludge age for nitrification using the aerated solids fraction only. Next, the anoxic zones are sized using denitrification rates appropriate for mixed liquor and sewage for the first anoxic zone and for mixed liquor endogenous respiration for the second anoxic zone. Typically, a 4:1 mixed liquor recycle is used between the first aerobic zone and the first anoxic zone. The size of the anaerobic zone is based on RAS flow and phosphate release rate measurements.

Both design procedures result in basin retention times in the range presented in Table 1. Basin retention times for the two full-scale Bardenpho plants in North America are also presented in Table 1.

TABLE 1 FIVE-STAGE BARDENPHO BASIN HYDRAULIC RETENTION TIMES

	Hydraulic Retention Time (h)		
Basin	Typical Range	Palmetto, Fl.	Kelowna, B.C.*
Anaerobic	1 - 2	1	2.9
First Anoxic	2 - 4	2.7	2.9
First Aerobic	3 - 8	4.7	8.6
Second Anoxic	2 - 4	2.2	3.8
Second Aerobic	0.5 - 1	1.0	1.9
Total	8.5 - 19	11.6	20.1

* Retention times are flexible by change of basin volumes devoted to each type of environment.

The process (with nitrification) has been reported to be limited by the COD/TKN ratio of the waste to be treated biologically. This limitation stems from the necessity to have enough soluble, readily metabolizable organic matter, both to stimulate anaerobic polyphosphate hydrolysis and to provide for denitrification. Limiting COD/TKN ratios of $\geq$ 12/1 have been quoted (Marais and Ekama, 1982). Recent data from Kelowna,

B.C. (Barnard et al., 1984; Oldham, 1984) seem to dispute this. However, Oldham acknowledges that the operation at Kelowna with a total COD/TKN ratio of as low as 8-9:1 is only successful because primary sludge gravity thickener overflow, containing readily metabolizable volatile fatty acids, is recycled to the anaerobic reaction zone.

UCT. Design procedures for the UCT process are similar to the semi-empirical biokinetic models used in the Bardenpho designs. It has been reported (Ikama et al., 1984) that EBPR in the UCT process can be maintained at a COD/TKN ratio as low as 7.5, compared to a minimum ratio of 12 for the Bardenpho process. As there is only one known full-scale example of the UCT process (Paepcke, 1985), full-scale confirmation of design and performance data are presently lacking.

A/O and A^2/O. Ranges of design criteria for the A/O process are presented in Table 2 (Hong et al., 1981). Choices of actual design values for specific applications, from within these ranges, are made by using a semi-empirical bio-kinetic model developed aby Air Products and Chemicals Inc. (Hong et al., 1981).

TABLE 2 TYPICAL A/O AND A^2/O DESIGN AND OPERATING PARAMETERS

Variable	Units	Range
Influent Retention Time		
Anaerobic Section	hours	0.5 - 1.0
Anoxic Section	hours	0.5 - 1.0
Aerobic Section		
Non-nitrifying (A/O)	hours	1.8 - 2.5
Nitrifying (A^2/O)	hours	3.5 - 6.0
F/M Ratio	kg BOD_5/day·kg MLVSS	0.15 - 0.7
Soluble BOD_5/Soluble P (Influent)	-	$\geq$10
Mixed Liquor Suspended Solids	mg/L	2000 - 4000
Temperature	°C	5 - 30
RAS Recycle Rate	% of Influent Flow	25 - 75
Internal Mixed Liquor Recycle Rate	% of Influent Flow	50 - 250
Basin Configuration		
Type	-	Staged System
Number of Stages:		
Anaerobic/Anoxic/Aerobic	-	3/3/4

A value of $\geq$ 10/1 for BOD_5/soluble P is required for the waste stream entering the biological portion of the A/O process. Staging, or compartmentalization of both the anaerobic and aerobic basins is recommended with at least three stages in the anaerobic and four stages in the aerobic basin. For the A^2/O process, a three-stage anoxic basin is added following the anaerobic basin. A mixed liquor recycle from the aerobic basin to the anoxic basin equivalent to 100% of the waste flow usually is sufficient for denitrification. For successful and virtually complete nitrification and denitrification, the influent waste to the A^2/O system should have COD/TKN $\geq$ 12. Each of the anaerobic and anoxic basins should be mixed. The oxygen requirements of the aerobic basin can be supplied by aeration or by the addition of high purity oxygen.

Phostrip. Phostrip process design parameters are required for three units: the P-stripper, the lime precipitator and the lime sludge clarification unit. Only the P-stripper design, which is unique to this process, is described.

Pierano (1977) developed empirical relationships between the efficiency of soluble orthophosphate removal (E%) by a Phostrip pilot plant employing a recycle P-stripper and various P-stripper parameters as follows:

$$E(\%) = 100 - 130^{(1.85 - P)}$$

where $$P = (120\ SL \times D)^{0.5} \times SU$$

and SL = sludge solids load on stripper, kg/ML

D = sludge solids retention time, h

SU = ratio of stripper supernatant flow to wastewater flow

More recently used P-stripper design criteria recognize the operation of the stripper as a sludge thickener and a device in which phosphorus is released from sludge solids. Miyamoto-Mills et al. (1983) suggested the following:

Solids retention time	$\geq$ 8 - 10 h
Elutriation rate	$\geq$ 0.3 l/g SS
Phosphorus release	$\leq$ 0.01 g P/g SS
Solids loading rate	- determined by RAS settleability but in the same range as for secondary activated sludge clarifier

Recycle rate (for recycle stripper) = 2-3 times sludge feed rate

These design criteria, developed for a Phostrip pilot plant, are in good agreement with those stated in the Biospherics Inc. Phostrip brochure (Anon, undated) as follows:

Solids retention time	8–12 h for 50% of RAS
Phosphorus release	0.005–0.2 g P/g VSS
Stripper depth	approx. 6.0 m
Stripper sludge blanket depth	approx. 4.8 m

2.1.5 **Process Performance Data.** Several detailed performance reviews of EBPR in pilot plant and full-scale installations have been made (Paepcke, 1983; Irvine et al., 1982; Walsh et al., 1983; Lamb, 1984; Paepcke, 1985). The information presented here is in summary form and has been updated with information generated since the completion of these reviews.

Paepcke (1985) reviewed available operational data to 1981 for 11 South African wastewater treatment plants reported to practice EBPR. His conclusions were that, out of the 11 plants, only one consistently achieved phosphorus removals to 1 mg/L orthophosphate. The remaining plants had highly variable EBPR histories. He acknowledged that the process limitations impeding a high level of biological phosphorus removal have been identified and measures to correct the situation were being implemented. This is not unusual and is to be expected in the development of new process technology.

Among other things, more recent process development work in South Africa has led to the use of the fermentation products from the acid digestion of primary sludge for EBPR process stabilization. Further, in order to ensure that a 1 mg/L orthophosphate phosphorus target will be achieved at South African wastewater treatment plants, chemical feed facilities to trim the effluent phosphorus after EBPR are being installed (Barnard, 1985). This approach follows a strategy of removing most of the phosphorus biologically and trimming it chemically. It is important to differentiate between the South African effluent phosphorus requirements which are expressed as orthophosphate and North American requirements, expressed as total phosphorus.

Tables 3, 4 and 5 present a summary of the status of Phostrip, Bardenpho and A/O plants derived from the data of Walsch et al. (1983), Lamb (1984) and recent telephone contacts. Table 6 presents a summary of North American full-scale EBPR plants.

As a result of their 1982 survey, Walsh et al. (1983) estimated the average effluent total phosphorus that EBPR plants could be expected to achieve (Table 7). On the basis of this survey they indicated that Kelowna, B.C. was the only EBPR plant meeting its effluent criteria without chemical addition beyond that anticipated in the

TABLE 3 SUMMARY OF PHOSTRIP PERFORMANCE (NORTH AMERICA)

Location	Design Flow (ML/d)	EBPR Flow Dec. 1984 (ML/d)	EBPR Start-up Date	EBPR Process Effluent Total P (mg/L)			Comments
				Standard	Average	Typical Range	
Seneca Falls, NY	6.7	0	1973	1.0	0.55	0.2 - 1.0	- 2 month test - Shutdown 1973
Carpentersville, IL	19	0	1979	Not Available (NA)			- Abandoned 1980; phosphate effluent standard not required
Lititz, PA	13	0	May 1982	-	-	-	- Low flows make EBPR inoperable - Chemical phosphate removal practised
Southtowns, NY	61	0	1983	1.0	1.2	NA	- Not in operation - Unable to obtain P release in stripper; low influent P
Texas City, TX	30	0	1983	2.0	NA	NA	- Not in operation - Unable to control stripper blanket
Amherst, NY	91	0	1981	1.0	-	0.7 - 1.7	- Shutdown March, 1983 - Ferric or ferrous iron addition more economical for the 3-4 mg P/L influent P
Lansdale, PA	9.5	5.7	April 1981	0.2 (Soluble)	0.7	0.5 - 1.0	- Have experimented with alum, but now returned to Phostrip
Adrian, MI	21	19	June 1981	1.0	-	0.1 - 1.0	- Current influent P is approx. 11 mg/L
Little Patuxent, MD Howard Co.	57	32	1981	0.3	2.0	1.0 - 2.0 (prior to alum addition)	- P standard achieved by combination of Phostrip and alum or sodium aluminate addition - Process is "touchy"
Reno Sparks, NV	114	87	Nov. 1981	0.7		0.5 - 6.0	- Alum added at 45-60 mg/L currently to secondary clarifier surface - Alum competes with Phostrip for P, interfering with P release in P-stripper
Tahoe Truckee, CA	21	21	Feb. 1984	0.8 (monthly avg.) 1.5 (daily avg.)	1.35 (0.52)* 3.65 (0.91)*	0.3 - 3.0 (0.31 - 0.67)* 0.8 - 8.4 (0.66 - 1.1)*	- Sensitive to high load and flow fluctuations. Alum added to primary clarifier to aid in combatting this.
Rochester, MN	73	73	July 1984	1.0	0.88	0.2 - 2.8	- High effluent SS (> 20-30 mg/L) require alum addition to secondary clarifier to meet < 1 mg P/L in effluent

* Phostrip effluent is blended with lime treated stripper supernatant then powdered and filtered through clinoptilolite. After filtration, values are given in parentheses.

TABLE 4 SUMMARY OF BARDENPHO PERFORMANCE (NORTH AMERICA)

Location	Design Flow (ML/d)	EBPR Flow Dec. 1984 (ML/d)	EBPR Start-up Date	EBPR Process Effluent Total P (mg/L)			Comments
				Standard	Average	Typical Range	
Palmetto, FL	5.3	4.6	October 1979	1.0	2.9	2-4	- Alum addition required to achieve P standards.
Kelowna, BC	22.8	11.4	June 1982	2.0	0.75	94% of Samples < 2 mg/L; 60% of Samples < 1 mg/L	- Primary sludge thickener overflow addition required to achieve EBPR. Better summer than winter performance.

TABLE 5 SUMMARY OF A/O PERFORMANCE (NORTH AMERICA)

Location	Design Flow (ML/d)	EBPR Flow Dec. 1984 (ML/d)	EBPR Start-up Date	EBPR Process Effluent Total P (mg/L)		Comments
				Standard	Typical Range	
Largo, FL	57	30-34	1 module 1981 3 modules Nov. 1983	1.0	2-3	- Effluent filters do not reduce total P significantly. Effluent P standard currently not enforced.
Pontiac, MI	5.3 - 7.6	4.9	March 1984	1.0	0.2 - 1.0	- Demonstration project in parallel with air activated sludge.
Pilot Plants						
Patapsco, MD	18 L/min	-	-	2.0	0.3 - 5.0	
Fayetteville, AK						- In start-up and preliminary phases of operation.

TABLE 6 PROCESS OPTIONS FOR CURRENTLY OPERATING EBPR PLANTS IN NORTH AMERICA

Location	EBPR Method	Flow Equal-ization	Liquid Stream Pro esses	Sludge Treatment Processes
Lansdale, PA	Phostrip	Yes	Act. Sludge + P-strip Fixed Film Nitrification Cl_2 Post aeration	- Lime stabilization, gravity thickening, land spreading
Adrian, MI	Phostrip	Yes	Prim. Sed. Act. Sludge + P-strip Act. Sludge Nitrification Filtration Cl_2	
Little Patuxent, MD	Phostrip	Yes	Prim. Sed. Act. Sludge + P-strip Act. Sludge Nitrification Filtration Cl_2 de-Cl_2 Reaeration	
Reno Sparks, NV	Phostrip	No	Prim. Sed. Act. Sludge + P-strip Reaeration Cl_2 de-Cl_2	- Phostrip lime sludge cosettled in prim. sed. - Anaerobic digestion, digested sludge lagoons
Tahoe Truckee, CA	Phostrip	No	Prim. Sed. Act. Sludge (O_2) + P-strip Clinoptilolite Filters	- Phostrip lime sludge settled separately, anaerobic digestion with biological sludge - Filter press + lime for digested sludge
Rochester, MN	Phostrip	No	Prim. Sed. Act. Sludge (O_2) + P-strip Act. Sludge (O_2) Nitrification Seasonal Cl_2	- Phostrip lime sludge settled separately and applied to land - Other sludges anaerobically digested, centrifuged and hauled away
Palmetto, FL	Bardenpho	No	Prim. Sed. (modified to fermentation) Act. Sludge + Bardenpho Filtration	- Drying beds, land disposal
Kelowna, B.C.	Bardenpho	Yes	Prim. Sed. Act. Sludge + Bardenpho Filtration Cl_2	- Prim. sludge gravity thickener - WAS, DAF thickener - Hauling for land disposal
Largo, FL	A/O	No	Prim. Sed. Act. Sludge + A/O	- Aerobic digestion, gravity thickening - Belt Press - Dry & Pelletize
Pontiac, MI	A/O	No	Prim. Sed. Act. Sludge + A/O	- Demonstration includes anaerobic digestion with supernatant return

TABLE 7 ESTIMATED EFFLUENT TOTAL PHOSPHORUS FROM EBPR PROCESSES*

Process		Achievable Effluent TP (mg/L)	
		Without Filtration	With Filtration
A/O	No nitrification	<2.0	<1.0
	With nitrification	<2.0	<1.5
	With nitrification & denitrification	<3.0	<1.5
Bardenpho		<1.0	<1.0
Phostrip		<1.5	<1.0

* Based on conclusions of Walsh et al. (1983).

process design. Lamb (1984) suggested that as of mid-1984, Adrian, MI, and Lansdale, PA, should be included in this category.

Lamb (1984) concluded that EBPR processes often are capable of producing excellent phosphorus removals from domestic wastewater (effluent phosphorus concentrations of 0.5 to 1.0 mg/L) but that the current state-of-the-art in their design and operation did not allow one to ensure consistent effluent phosphorus concentrations of less than 1 mg/L.

Since the reviews by Walsh et al. (1983) and Lamb (1984), detailed operating data from two full-scale EBPR plants have become available. These are both Phostrip plants - the Tahoe Truckee Sanitation Agency (TTSA), CA, plant and the Rochester, MN, plant. The unit processes employed in these plants are noted in Table 6 and their performances given in Table 3. Since they represent the latest application of Phostrip technology, a more detailed presentation and discussion follows.

TTSA is a Phostrip retrofit to a plant that was oxygen activated sludge followed by high lime precipitation, recarbonation and clinoptilolite sorption for nitrogen removal. The high lime for phosphorus removal was replaced with Phostrip added to the RAS system of the oxygen activated sludge plant. The stripper overflow is precipitated with lime in one of the original high lime precipitation/sedimentation basins and then blended with the secondary effluent prior to feeding to an equalizing pond followed by clinoptilolite sorption. Table 8 gives average monthly and maximum daily total phosphorus concentrations for February to December, 1984. During this period, both start-up and experimentation occurred. Many of the experiments were designed to find

methods for ameliorating the effects of fluctuating flows and loads on the Phostrip process. These included feeding raw sludge to the secondary system or removing primary clarifiers from service (to combat dilute sewage); flow equalization in an emergency holding basin (to combat fluctuating flows); adding alum to primary clarifier to reduce phosphorus by approximately 1 to 2 mg/L (to combat fluctuating load).

TABLE 8 30-DAY AVERAGE AND DAILY MAXIMUM TOTAL P CONCENTRATIONS AT THE TAHOE TRUCKEE SANITATION AGENCY, CA, PHOSTRIP PLANT

	Total Phosphorus (mg/L)						
	Influent		Secondary Effluent		Filtered Effluent		Average Monthly Effluent Temperature (°C)
Date 1984	30-day Avg.	max. day	30-day avg.	max. day	30-day avg.	max. day	
February	7.0	10.6	1.2	2.2	0.57	1.0	7
March	10.3	11.5	1.9	2.8	0.64	1.1	8
April	5.6	7.2	0.82	1.9	0.48	0.72	8
May	5.8	9.0	0.30	0.78	0.31	0.66	10
June	6.9	8.8	1.4	4.9	0.47	0.73	13
July	8.3	9.1	3.0	8.4	0.67	0.98	17
August	9.1	10.4	1.2	3.6	0.63	1.0	18
September	7.7	9.9	1.8	5.3	0.60	0.91	17
October	7.0	9.2	1.5	3.2	0.67	1.1	14
November	6.4	9.3	1.2	5.6	0.34	0.88	11
December	7.1	10.4	0.53	1.5	0.34	0.93	8

After filtration (clinoptilolite sorption), the effluent phosphorus criteria (30-day average = 0.8 mg P/L; daily maximum = 1.5 mg P/L) were met (30-day average = 0.31 -0.67 mg P/L; daily maximum = 0.66 - 1.1 mg P/L). The activated sludge plus Phostrip, unfiltered and unblended with lime precipitated and settled stripper supernatant did not always meet the criteria (i.e., 30-day average 0.3 - 3.0 mg P/L; daily maximum = 0.8 -8.4 mg P/L).

The Rochester, MN, plant is a two-stage oxygen activated sludge plant, designed for carbonaceous BOD removal in the first stage and nitrification in the second

stage. It is in conjunction with the first stage that the Phostrip process is used. The Phostrip process had been in operation since July 1984; the data for July to December, 1984 are summarized in Table 9. It was found necessary to add from 2725 to 5450 L of alum per day to the final clarifiers to meet effluent suspended solids and phosphorus standards (Skodje, 1985). It was not known whether this was due to the centrate being returned to the plant or because of high sludge age. It may also have been due to inexperience in Phostrip operation.

TABLE 9 30-DAY AVERAGE AND DAILY MAXIMUM EFFLUENT TOTAL P AND SS CONCENTRATIONS AT THE ROCHESTER, MN, PHOSTRIP PLANT

Date 1984	Effluent Total Phosphorus (mg/L)		Effluent Total Suspended Solids (mg/L)	
	30-day avg.	daily max.	30-day avg.	daily max.
July	0.7	1.1	15	38
August	0.8	1.1	21	46
September	0.9	1.4	18	34
October	1.0	1.4	19	36
November	1.0	2.8	14	33
December	0.9	1.9	18	36

In addition to the full-scale Phostrip plants at TTSA and Rochester, a prototype EBPR plant has been operated since the time of the reviews by Walsh et al. (1983) and Lamb (1984). This is a demonstration of the A/O process at Pontiac, MI. At Pontiac, one half of an existing 14.6 ML/d air activated sludge plant has been converted to the A/O configuration and is being run side-by-side with air activated sludge. The purpose of the project is to investigate A/O phosphorus removal performance at low temperature under nitrifying and non-nitrifying conditions and to assess the impact of anaerobic digester supernatant recycle on plant performance. Detailed data will not be released until the conclusion of the study in May 1985. An Air Products and Chemicals Inc. news release dated October 1984 presents the limited information in Table 10.

A pilot plant study of the A/O process was recently started in Fayetteville, AK (December, 1984). For the very limited period of operation, the pilot plant produced an effluent with an average total phosphorus of 0.3 mg/L from an influent with an average total phosphorus of 1.9 mg/L.

TABLE 10 SUMMARY OF PONTIAC, MI, A/O PROCESS PERFORMANCE

Period	Temperature (°C)	Diurnal Flow Fluctuations	Parameter	Influent (mg/L)	Effluent (mg/L)	
					A/O	Conventional
May - Sept. 1984	13-23	Yes	BOD_T	118	8	7
			TSS	61	7	8
			TP	3.4	0.4	1.5
			NH_3-N	15.5	1.0	1.0

2.1.6 Sludge Generation Data. There is no field evidence to suggest that any more activated sludge is produced by EBPR than by activated sludge removing "normal" amounts of phosphorus at the same growth rate. Indeed, Air Products and Chemicals Inc. claims that the A/O Process produces less activated sludge than conventional activated sludge systems. In support of this claim, it is stated that for the Patapsco pilot plant (Deakyne et al., 1984), the sludge production was 0.6 to 0.7 kg TSS/kg BOD_5 removed, whereas the design value for the conventional activated sludge system was 0.83 kg TSS/kg BOD_5 removed. However, a sludge production of 0.6 to 0.7 kg TSS/kg BOD_5 removed is well within the typical range of sludge production for conventional activated sludge treating domestic wastewater.

The sludge production by the Phostrip process will be somewhat greater than for the A/O and Bardenpho processes because a chemical sludge is produced from the lime precipitation of stripper supernatant. For example, at the TTSA plant, dry solids lime sludge production alone is 126 kg/1000 m^3 influent at an effluent total phosphorus of 0.6 mg/L. This is over and above the primary plus biological sludge dry solids production.

2.1.7 Process Sensitivity and Response to Low Temperature. The available data and experience seems to indicate that EBPR performs better at low temperature (e.g., 5°C) than at higher temperatures. Sell et al. (1981) conducted sequencing batch laboratory scale experiments on the A/O process treating a synthetic waste (BOD_5 = 75 mg/L; TP = 5 mg/L) under conditions where nitrification was inhibited and the temperature ranged from 5 to 15°C. EBPR was not compromised at the lower temperature as long as the F/M ratio did not exceed 0.25 kg BOD_5/kg VSS-day; in fact, phosphorus removal rates (kg P/kg VSS-h) were higher at 5°C than at 15°C. Sell et al. theorized that their results could be explained if it was assumed that the polyphosphate accumulating microorganisms responsible for EBPR were psycrophilic.

The operator of the TTSA Phostrip plant reported that the plant "worked better" at low temperatures, although the operating data do not substantiate this comment. However, it is likely that this observation was influenced by many things other than temperature. The average monthly effluent temperature at TTSA ranged between 7°C and 18°C.

Evaluation of the low temperature performance of the A/O process is one of the objectives of the full-scale demonstration study underway at Pontiac, MI. No detailed data on temperature effects were available.

From the available data, it can be tentatively concluded that for non-nitrifying Phostrip and A/O processes, temperatures as low as 5°C can be accommodated without compromising the ability of EBPR. This conclusion can be stated in terms relative to other biological treatment objectives by saying that, at low temperature, nitrification rather than EBPR will be the rate limiting biological process.

2.1.8 Process Response to Organic, Toxic and Hydraulic Shock and Upset Conditions. Table 6 shows that many of the functioning EBPR plants have some type of flow equalization. Further, the operating data produced from the A/O process at Largo, FL, when used in only one module of the plant, was from a constant influent flow. Now that the entire Largo plant has been converted to the A/O process, it is subject to normal diurnal fluctuations in flow and load.

In general, EBPR processes seem able to cope with modest load and flow variations. For example, the A/O pilot plant at Patapsco was subjected to flow variations in two ranges: 79 to 130% of average and 104 to 173% of average, representing peak to average flow variation of 1.3:1 for dry and 1.7:1 for wet weather conditions, respectively. The results (Table 11) show that the process monthly average effluent total phosphorus was 1.4 mg/L at the lower flows and 2.2 mg/L at the higher flows.

TABLE 11 A/O PILOT PLANT PERFORMANCE AT PATAPSCO UNDER INFLUENT FLOW VARIATION

Flow Range (% of Average)	Effluent P (mg P/L)		
	Average Monthly		Weekly Maximum
	Soluble	Total	Total
79 - 130	0.64	1.4	1.6
104 - 173	1.65	2.2	3.0

It is with more extreme and sudden flow and load fluctuations that difficulties have been reported. For example, at TTSA, the influent waste is subject to large variations in strength and flow due to the influx of weekend skiers in the winter and infiltration/inflow in the spring. Under such conditions, excursions to high effluent phosphorus concentrations are experienced. Amelioration has been by:

i) flow equalization using an emergency retention basin (flow fluctuation),
ii) feeding raw sludge to the secondary system or removing the primary clarifier from service to feed raw sewage (low load), and
iii) adding alum to the primary clarifier to achieve a 1 to 2 mg P/L reduction (low load).

Start-up of EBPR processes appears to be slower than for secondary treatment process start-up. The Patapsco A/O pilot plant took approximately two months to achieve EBPR while secondary treatment plants usually can be started up to produce BOD removal in 10 to 14 days (somewhat longer if nitrification is incorporated). At Patapsco, the use of an activated sludge seed from an A/O plant actively achieving EBPR did not accelerate start-up. Once achieved at a particular plant, however, EBPR appears to be relatively amenable to restarting after shutdown. For example, the A/O plant at Largo achieved an effluent total phosphorus of less than 1 mg/L after having been shut down for a period of 14 days during which the A/O sludge was stirred but not aerated.

There seems to be no reason to assume that the recovery of EBPR plants from toxic slugs is any different from that of the BOD-removing processes. This is supported by observations at Patapsco. EBPR was inhibited by toxic discharges on two occasions, but the A/O process returned to the prior phosphorus removal performance within 24 hours on both occasions.

It would seem reasonable to assume that EBPR plants that also must nitrify (and therefore denitrify) would be more susceptible to flow and load variations than those that do not. In nitrifying plants, the BOD and nitrogen should be present in a ratio that allows satisfactory denitrification at all times.

2.1.9 Operating Problems. Operating problems can be divided into two groups: those associated with mechanical equipment and design, and those associated with process capability. In many ways, these are related because the ability of the process to cope with different conditions is a function of its flexibility of design.

Early Phostrip plants were plagued with a variety of mechanical problems, especially in lime feeding. Comments received from some operators also indicated the

absolute necessity for flexibility. For example, the provision of more than one P-stripper was cited as essential.

Flexibility, and also instrumentation for process control, appear to be key elements in the design of the Bardenpho process for nitrifying/denitrifying systems. The ability to "shuffle" basin volumes assigned to aerobic, anoxic and anaerobic regimes is valuable; the measurement of dissolved oxygen in aerated basins and oxidation-reduction potential (ORP) in anoxic and anaerobic basins, and the ability to use this to control aeration and mixing are important.

Fluctuating flows and loads have been identified as problems especially for nitrifying EBPR plants when the BOD/N and BOD/P ratios vary widely over the short term. Flow equalization is practiced in many of the full-scale EBPR plants to counteract these problems. In some cases, supplementary chemical addition is required.

The maintenance of anaerobic conditions in the initial contact tank is essential to efficient phosphorus release and the presence of dissolved oxygen and nitrate in the anaerobic basin may contribute to the less than optimum performance of many existing EBPR systems. The presence of either DO or nitrate will reduce the concentration of readily metabolizable COD available to the microorganisms for phosphorus release. Elimination of hydraulic re-aeration of the raw wastewater and control of recycled nitrate through complete denitrification are critical to process performance.

Sludge treatment processes have a strong impact on EBPR. Since anaerobic conditions favor phosphorus release from sludge that has taken up phosphorus by EBPR, the recycle of sludge streams from anaerobic sludge treatment processes will re-introduce phosphorus to the waste stream. For this reason, most of the EBPR processes do not recycle anaerobic digester supernatant to the main stream (Table 6). Activated sludge concentration prior to further treatment is by dissolved air flotation (DAF), an aerobic process. Sludge treatment following anaerobic digestion is by either removal from the plant site or by a concentration process that employs high pH (e.g., lime addition and filter pressing) so that the phosphorus released in the liquid stream from sludge digestion is precipitated in the sludge concentration step.

Where digester supernatant recycle is practiced, as in the typical Canadian plants removing phosphorus by chemical addition to a primary/secondary treatment plant, one of two process modifications will have to be adopted if EBPR is to be successfully practiced:

i) stop supernatant recycle; operate digester as mixed digester; dispose of mixed digester contents off-site or treat mixed digester contents by a sludge

concentration procedure that employs high pH, or iron or aluminum addition to precipitate phosphorus in the sludge solids, or

ii) continue supernatant recycle but treat chemically to precipitate phosphorus using iron or aluminum addition equipment.

2.1.10 Technical Feasibility of EBPR Processes. Based on recent experiences with full-scale EBPR plants, it can be concluded that the process capabilities as determined by Walsh et al. (1983) and Lamb (1984) are generally accurate. With EBPR processes alone, an effluent total phosphorus limit of 1.0 to 2.0 mg/L can be achieved if the practice of anaerobic digester supernatant recycle were terminated or if digester supernatant were treated chemically to remove released phosphorus prior to recycle to the biological process.

To achieve an effluent total phosphorus limit of 1.0 mg/L by the totally biological phosphorus removal processes (A/O, Bardenpho, UCT), effluent filtration will be required to reduce the particulate phosphorus contributed by the phosphorus-rich biological solids in the clarified effluent. Alternatively, chemical precipitants would need to be applied in conjunction with the EBPR process to reduce the residual soluble orthophosphate concentration to a lower level than typically achieved by EBPR alone. For the Phostrip process, full-scale operating experience indicates that an effluent total phosphorus limit of 1.0 mg/L can be achieved without effluent filtration or supplemental chemical addition to the biological system.

To consistently achieve an effluent total phosphorus limit of 0.3 mg/L, all of the EBPR processes (A/O, Bardenpho, UCT and Phostrip) would require final effluent filtration. In addition, A/O, Bardenpho and UCT would require supplemental chemical addition to reduce the soluble orthophosphate concentration to a low level (< 0.3 mg/L).

Table 12 summarizes the process configurations required to achieve a range of effluent total phosphorus concentration limits. While the process requirements indicated in Table 12 appear to be somewhat more stringent than those of Walsh et al. (1983) and Lamb (1984), they are meant to be all-inclusive in their application. The achievement of effluent total phosphorus values lower than those indicated may be possible by EBPR processes if several design features are included in the plant. Among these features are:

i) flow and load equalization to the EBPR process;

ii) exclusion of recycle streams containing high phosphate (soluble and/or particulate);

iii) compartmentalization of the EBPR basins;

TABLE 12 EBPR PROCESS REQUIREMENTS FOR VARIOUS EFFLUENT TOTAL PHOSPHORUS LIMITS

Effluent Total P Concentration (mg/L)	Process Configuration for			
	Bardenpho	UCT	A/O	Phostrip
1.0 - 2.0	EBPR	EBPR	EBPR	EBPR
1.0	EBPR and Filtration or EBPR and Chemical Addition	EBPR and Filtration or EBPR and Chemical Addition	EBPR and Filtration or EBPR and Chemical Addition	EBPR
0.3	EBPR and Filtration and Chemical Addition	EBPR and Filtration and Chemical Addition	EBPR and Filtration and Chemical Addition	EBPR and Filtration

iv) flexibility in use of EBPR basin volumes between anaerobic, anoxic, and aerobic environments;

v) ability to provide a source of readily metabolizable soluble carbon in the influent to the EBPR process; and

vi) maintenance of an oxygen-free and nitrate-free environment in the anaerobic basin by provision of denitrification of streams recycled to it and by avoiding aeration of the influent.

These features recognize the results of recent research that emphasizes the importance of "anaerobic" conditions in the anaerobic basin and the need for a readily metabolizable carbon source to ensure phosphorus release. Further, they recognize the field experience that large fluctuations in load and flow, together with a plant with little flexibility in basin environment and configuration, can lead to temporary excursions from the low effluent phosphorus concentrations.

In addition to these process design features, it appears that achieving low effluent total phosphorus limits is related to the influent phosphorus concentration. At the low influent phosphorus levels typical of areas where phosphorus limits in detergents are imposed, EBPR processes alone are more capable of meeting or exceeding a 1.0 mg/L total phosphorus requirement.

Several of the existing prototype EBPR plants incorporate flow or load equalization (Table 6). Those that do not, and have greater and more sustained flow and/or load changes than typical (e.g., 1.3 to 1.5 peak/average), appear to have difficulty in not producing excursions to higher than 2 mg/L total phosphorus in the effluent. For this reason, it would be desirable, in a retrofit situation, to keep the existing chemical addition system and a supply of chemicals on hand to use on occasions of EBPR process upset. This is practiced in several of the existing EBPR full-scale plants.

EBPR plants, like any biological treatment process, are susceptible to toxic shocks; however, the EBPR process does not seem to be any more susceptible than the BOD removal process. Recovery seems to be fairly rapid in the cases cited (24 to 48 h). Start-up of the EBPR process from non-EBPR sludge is rather slow, reported to be approximately two months in one case. Restarting the mechanism from a sludge that has recently been exhibiting EBPR is quite rapid, in the order of one to two days.

The EBPR process seems to function well in a temperature range of 5°C to 25°C. Some investigators suggest that the performance is better at lower temperatures. At least it can be stated that EBPR behaves as well as BOD_5 removal at these temperatures.

Since the phosphorus sink is either activated sludge suspended solids alone or in combination with a lime/phosphate sludge, anoxic conditions and low pH will cause release of the phosphorus from these solids back into solution. These conditions should be avoided in the mainstream of the plant and in any sidestreams returned to the mainstream. Sludge processing and disposal techniques are both critical and limited by EBPR. Activated sludge should be concentrated using "aerobic" processes, such as air flotation, rather than gravity thickening. Anaerobic digester supernatant return should be eliminated by operating mixed digesters or, if not eliminated, the supernatant must be treated with a chemical (e.g., lime) to precipitate phosphorus and then settled to remove the precipitated phosphorus prior to recycle. Sludge processing should incorporate chemicals for sludge conditioning (e.g., lime, $FeCl_3$) to ensure that the phosphorus ends up in the sludge phase rather than in the liquid phase that is returned to the plant mainstream. An example is the use of lime in a sludge filter press. Ideally, disposal of mixed digester sludge to land off-site can be used to eliminate all sludge processing return streams.

The EBPR plants designed and built most recently provide much more flexibility and control, and appear to function better and more reliably under less than optimum conditions than the earlier plants. It seems important to provide compartmentalization of the various basins and to have the flexibility to shuffle basins between different environments (anaerobic, anoxic, aerobic). The measurement and control of DO, and possibly ORP, are recommended. Multiple basins are desirable (e.g., two smaller P-strippers rather than one larger one for the Phostrip process).

2.1.11 Summary. EBPR can be a technically feasible component of a process that is required to produce effluent total phosphorus in the range from 0.3 to 2 mg P/L. The ease with which EBPR can fulfill this role is a function of whether nitrification, and necessary denitrification, is required; the influent phosphorus concentrations and influent BOD_5/N and BOD_5/P ratios; the variations in influent flow and load; the presence of toxic materials; and the nature of the sludge processing methods practiced or proposed.

In retrofitting EBPR processes to existing biological treatment plants or primary treatment plants that use chemical precipitation for phosphorus removal, it would be advisable to leave the chemical dosing facilities in place. They could be useful (or indeed may be required) in meeting the effluent total phosphorus criteria, and in coping with fluctuating loads or treating sludge processing recycle streams.

2.2 Chemical Phosphorus Removal

Until recently, chemical precipitation of phosphorus has been the only practical, proven technology for the reduction of phosphorus at municipal waste treatment plants that are required to meet effluent total phosphorus limits. Since chemical precipitation processes are easy to design, have low capital cost and are simple to start-up and operate, they are attractive from an engineering and operational perspective.

2.2.1 Process Alternatives. A number of chemical precipitation process options have been successfully used for the reduction of total phosphorus at municipal waste treatment plants. These range from pre-precipitation, to simultaneous precipitation, to post-precipitation. The choice of alternatives depends on many factors. In a process retrofit situation for a conventional activated sludge installation, phosphorus removal through simultaneous precipitation has been found to be a popular, simple and cost-effective solution in Ontario.

2.2.2 Process Descriptions. Pre-precipitation technology involves adding the precipitant at the primary clarification stage and settling the resulting chemical sludge together with the primary sludge. Generally, in this case, lime is used as the precipitant.

Simultaneous precipitation involves the addition of a metal salt, such as alum or ferric chloride, to the activated sludge aeration tank contents in order to precipitate the hydrolyzed phosphates. The resulting mixed biological and chemical sludges settle in the secondary clarifier. The waste activated sludge, combined with the chemical sludge, is stabilized either through anaerobic or aerobic digestion. In Ontario, simultaneous precipitation of phosphorus is practised at conventional activated sludge and extended aeration plants. Extensive bench, pilot and full-scale testing of this technology was conducted in response to the Ontario requirement to reduce effluent total phosphorus concentrations to 1 mg/L. The data base that was developed served to establish design criteria for retrofitting existing plants for simultaneous precipitation using alum or iron salts.

Post-precipitation technology is not widely practised in Ontario. It involves the construction of additional process units following the secondary clarifier. If metal salts are used, process units such as a reactor/flocculator and clarifiers are required. Past Scandinavian practice has been to use this approach for phosphorus removal. It should be noted that the Scandinavian total phosphorus effluent target was 0.5 mg/L. For

total phosphorus effluent concentrations of 0.1 mg/L and less, single or two-stage lime systems with recarbonation are used.

The addition of chemicals for phosphorus removal at Ontario plants is on a continuous basis. In the past, chemical addition to Ontario waste stabilization ponds with seasonal discharge was practiced on a batch basis, approximately two weeks before discharge. Experience has shown that, based on practical considerations, continuous feeding is preferred.

2.2.3 Process Design Criteria. One of the many process development and assessment studies conducted under the Canada/Ontario phosphorus control program examined the impact of chemical addition to the aeration tank on the settleability of the combined biological-chemical sludge. It was observed that sludge settleability decreased. Consequently, secondary clarifier overflow rates were reduced from 50 m^3/m^2·day to 35 m^3/m^2·day at peak flow (IJC, Municipal Task Force, 1983). This ensured that the effluent suspended solids concentration remained at less than 15 mg/L, and the total phosphorus at no greater than 1 mg/L. A fraction of the phosphorus is, of course, associated with the suspended solids.

Insofar as the choice of chemical precipitant is concerned, both alum and ferric chloride are used in Ontario. Generally, either precipitant is equally effective. The final choice usually depends on the cost and availability of the precipitant. Chemical dosage requirements were initially based on jar tests and fine-tuned in the field. The data base developed in Ontario now enables the design engineer to predict the chemical dosage required based on the relationship (Prested et al., 1977):

$$Al^{+++}\ (mg/L) = 0.5 + 1.4\ (TP)$$

$$Fe^{+++}\ (mg/L) = -0.3 + 1.3\ (TP)$$

where: Al^{+++} or Fe^{+++} = precipitant metal dosage (mg/L)

TP = influent TP concentration (mg/L)

It has been found that these equations predict the actual chemical dosage to within $\pm$ 30%. This degree of accuracy is adequate to design chemical storage and feeding equipment.

It has been observed that at Ontario treatment plants, the average influent total phosphorus is 5.4 mg/L. The average chemical dosage to attain an effluent total phosphorus concentration of 1.0 mg/L is 9.2 mg/L for ferric chloride and 7.2 mg/L for alum (Canviro, 1983).

2.2.4 Process Performance Data. The extensive Ontario experience, which has been previously reported (Black, 1980; Schmidtke, 1980), has demonstrated that simultaneous precipitation of phosphorus using metal salts is an effective approach to controlling the discharges of phosphorus from municipal wastewater treatment plants. Some 272 Ontario plants practice phosphorus removal (Schmidtke, 1983). The flow-weighted average effluent total phosphorus concentration is 1.21 mg/L (IJC, Municipal Task Force, 1983). There are also plants in environmentally sensitive areas where the effluent total phosphorus concentration must be 0.5 mg/L or even 0.3 mg/L. There is no difficulty in achieving this requirement with a well functioning biological process using simultaneous precipitation followed by effluent filtration through granular media filters. Specific sites may require a small additional precipitant dosage just prior to filtration.

2.2.5 Sludge Generation Data. The addition of chemicals for phosphorus removal results in increased sludge generation. The Ontario experience using metal salts has been summarized by Schmidtke (1980) as shown in Table 13.

TABLE 13 EFFECT OF CHEMICAL PHOSPHORUS REMOVAL ON SLUDGE GENERATION RATES

Plant Type	Operating Mode	Sludge Quantity		
		g/person-d*	% Volume of Effluent	kg Dry Solids per 1000 m^3 Treated
Primary	Conventional	77	0.20	120
	P Removal	109	0.32	169
Activated Sludge	Conventional	114**	0.38	173
	P Removal	145**	0.51	218

* Based on Q = 658 L/person-d
** Primary + Waste Activated

The Ontario experience has been that sludge production at primary plants using metal salts for phosphorus precipitation from 5.7 mg/L to 1.0 mg/L results in sludge mass increases of as much as 40%. Similarly, sludge mass may increase by as much as 26% for conventional activated sludge plants (Schmidtke, 1980).

2.2.6 Process Sensitivity, Response and Recovery from Upset Conditions. Phosphorus removal using chemical precipitation and biological treatment simultaneously

can be adversely affected if the chemical dosage is insufficient to react with the phosphorus. This may be triggered by malfunctioning chemical dosing equipment, high influent phosphorus concentrations and compounds in the influent that compete with phosphorus for the metal precipitant. As well, hydraulic load shocks which adversely affect liquid-solid separation will result in phosphorus carryover into the plant process effluent. Any of these conditions will result in a quick, adverse process response so that effluent total phosphorus concentrations will be greater than design.

In the context of municipal waste treatment technology, chemical processes exhibit more stability than biological processes. If a chemical process is used in conjunction with a biological process, such as simultaneous precipitation, then the overall process would benefit from the stabilizing influence of the chemical process.

Recovery from 'upset' conditions will occur if the cause is rectified. Remedial procedures are generally mechanical or structural. Once rectified, the process response is immediate and positive.

2.2.7 Operating Problems. As noted in the previous section, operating problems at treatment plants have resulted in higher than targeted effluent total phosphorus concentrations. These problems are attributable to hydraulic overloading of the final clarifier caused by such events as combined sewer overflows and infiltation, and high solids carryover in the effluent due to the inability of clarifiers to process the solids going to the clarifiers. Other problems have been minor and are to be expected in any operation using chemical feeders and pumps.

2.2.8 Technical Feasibility. Chemical precipitation technology for phosphorus removal is well proven in Canada and elsewhere. The Ontario experience has demonstrated that the technology, when used in the simultaneous precipitation mode, will consistently produce effluent total phosphorus concentrations of 1.0 mg/L if adequate clarification capacity is provided. Effluent total phosphorus concentrations to 0.3 mg/L can be produced by effluent filtration after chemical addition.

3 STUDY METHODOLOGY

The comparative costs for retrofitting existing primary and secondary (conventional activated sludge and extended aeration) plants with EBPR processes and traditional chemical precipitation technologies were developed using the methodology described in this section. This cost analysis is based on 'model plants' having idealized configurations, although standard design practices were followed as much as possible. The retrofit costs of the biological nutrient control technologies will be influenced by the actual plant equipment and the plant layout at a specific location.

3.1 Design of Model Plants

Three types of plants were considered for retrofitting to EBPR processes, i.e., primary plants, conventional activated sludge plants and extended aeration plants. A standardized design for each type of plant was developed, based on Ontario Ministry of the Environment design guidelines (1984). The resulting standardized plant configurations are referred to as 'model plants'. These model plants formed the basis for the process retrofits to achieve phosphorus control.

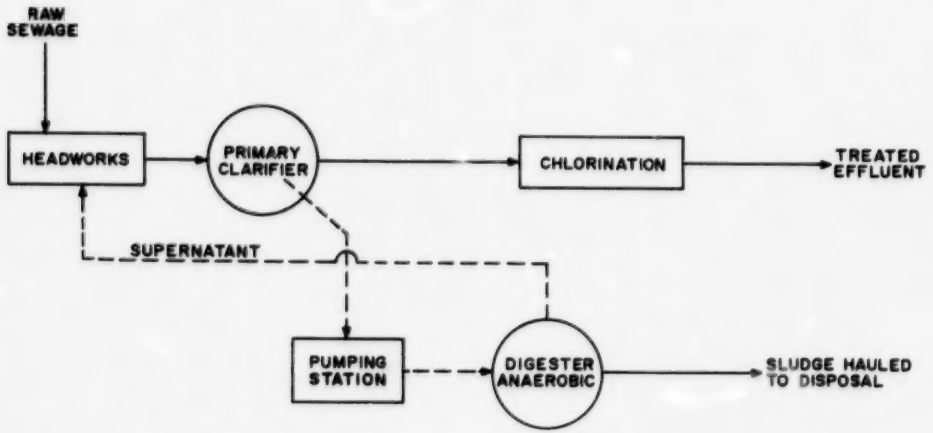
3.1.1 Model Primary Plant. The process flowsheet for the model primary plant is shown in Figure 3. The plant consisted of headworks (bar screen, aerated grit chamber, comminutor and Parshall flume), primary clarifier including raw sludge pumping station, single-stage anaerobic digester and chlorine contact chamber. Design criteria for the individual units are summarized in Figure 3.

3.1.2 Model Conventional Activated Sludge Plant. The process flowsheet for the model conventional activated sludge plant is shown in Figure 4. The plant incorporated the same headworks as the model primary plant, primary clarifiers with raw sludge pumping station, aeration basins with mechanical surface aerators, secondary clarifiers with return sludge pumping station, two-stage (primary and secondary) anaerobic digestion and chlorine contact chamber. Design criteria for the individual units are summarized in Figure 4.

3.1.3 Model Extended Aeration Plant. The layout of the model extended aeration plant deviated from standard design practices. In the smaller size ranges (up to 1.0 to 2.0 ML/d), packaged treatment plant configurations are commonly used. These small packaged treatment plants, which typically incorporate common-wall, steel-tank construction for aeration tanks, clarifier and sludge holding tanks, are not well-suited for

PLANT TYPE : PRIMARY

PROCESS FLOW SHEET



Design Criteria

Plant Hydraulic Capacity (ML/d)

25% tile	-	2.3
50% tile	-	9.1
75% tile	-	36.4

Headworks:

- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

Clarifier:

Surface Overflow Rate	
@ Peak Flow ($m^3/m^2 \cdot d$)	35-40
Weir Overflow Rate	
@ Peak Flow ($m^3/m \cdot d$)	150-450
Sidewater Depth (m)	3.0-4.6

Anaerobic Digestion:

Volatile Solids Loading	
(kg $VS/m^3 \cdot d$)	0.65-1.6
Volatile Solids Destroyed (%)	50

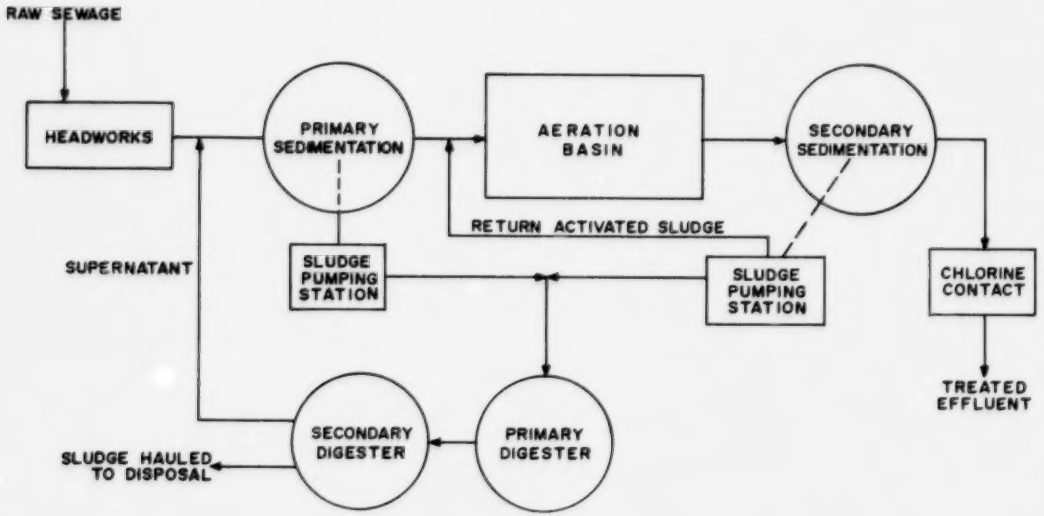
Chlorine Contact:

- 30 min HRT
- 8-15 mg/L Chlorine Dosage

FIGURE 3 MODEL PRIMARY PLANT PROCESS FLOWSHEET

PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

PROCESS FLOW SHEET



Design Criteria

Plant Hydraulic Capacity (ML/d)

25% tile	-	4.5
50% tile	-	13.6
75% tile	-	36.4

Headworks:
- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

Primary Clarifier:

Surface Overflow Rate	
@ Peak Flow ($m^3/m^2 \cdot d$)	80-120
Sidewater Depth (m)	3.0-4.6

Aeration System:

Organic Loading Rate	
(g BOD/$m^3 \cdot h$)	13-30
F/M (d^{-1})	0.2-0.5

Secondary Clarifier:

Surface Overflow Rate	
@ Peak Flow ($m^3/m^2 \cdot d$)	35
Sidewater Depth (m)	3.6-4.6

Anaerobic Digestion:

Volatile Solids Loading	
(kg VS/$m^3 \cdot d$)	0.65-1.6
Volatile Solids Destroyed (%)	50

Chlorine Contact:
- 30 min HRT
- 8-15 mg/L Chlorine Dosage

FIGURE 4 MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT PROCESS FLOWSHEET

retrofitting to biological phosphorus removal processes. Therefore, the extended aeration plants were designed as standard mechanical plants with separate components rather than as packaged plants for the purposes of this cost assessment.

The process flowsheet for the model extended aeration plant is shown in Figure 5. Process components included the headworks as in the other model plants, the aeration basin equipped with mechanical aerators, secondary clarifier with return sludge pumping station, an aerated sludge holding tank and the chlorine contact chamber. Design criteria were as summarized in Figure 5.

3.1.4 Size of Model Plants. Plant sizes for each category of model plant were established based on the distribution of actual sizes of plants in Ontario in each category. Frequency distributions of design flows for primary plants, conventional activated sludge plants and extended aeration plants in Ontario were developed. The 25 percentile, 50 percentile and 75 percentile sizes in each category were selected as the basis for the model plants. The model plant sizes are summarized in Table 14.

TABLE 14 DISTRIBUTION OF TREATMENT PLANTS IN ONTARIO ACCORDING TO TYPE AND SIZE

Plant Type	Number of Plants	Hydraulic Flow (ML/d) 25 Percentile	50 Percentile	75 Percentile
Primary	33	2.3	9.1	36.4
Conventional Activated Sludge	88	4.5	13.6	36.4
Extended Aeration	85	0.91	1.6	3.2

Since the hydraulic loading to primary and secondary clarifiers was based on peak flow, peaking factors were calculated using the Harmon formula:

$$PF = 1 + \frac{14}{4 + p^{0.5}}$$

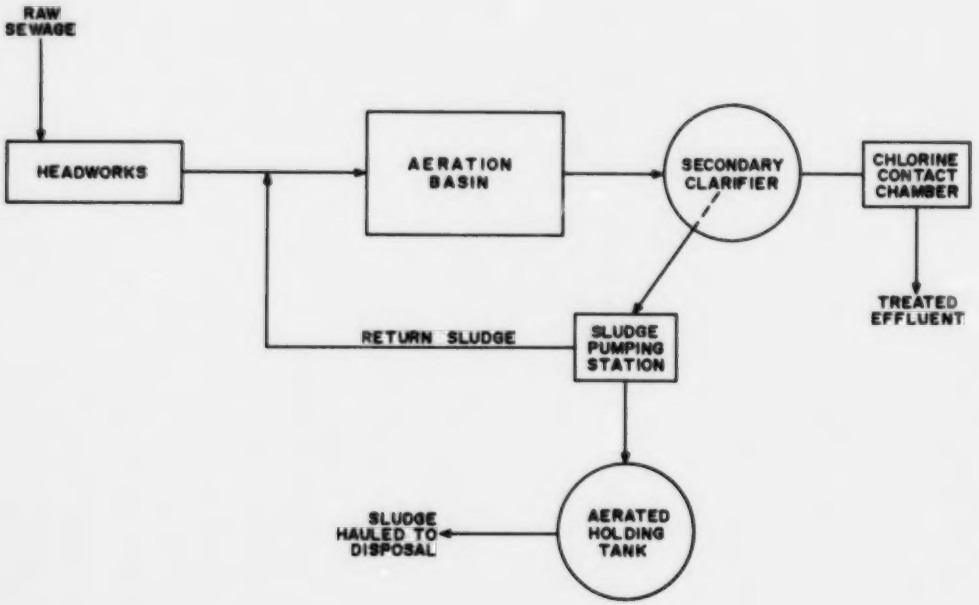
where: PF = peaking factor

P = population in thousands

In accordance with Ontario data, design populations for the various flows were calculated using an average per capita wastewater production value of 650 litres per day.

PLANT TYPE : EXTENDED AERATION ACTIVATED SLUDGE

PROCESS FLOW SHEET



Design Criteria

Plant Hydraulic Capacity (ML/d)

25% tile	-	0.91
50% tile	-	1.6
75% tile	-	3.2

Headworks:

- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

Aeration System:

Organic Loading Rate	
@ Peak Flow (g BOD/m^3·h)	7-10
F/M (d^{-1})	0.05-0.15

Secondary Clarifier:

Surface Overflow Rate	
@ Peak Flow (m^3/m^2·d)	35
Sidewater Depth (m)	3.0-4.6

Aerobic Digestion:

Volatile Solids Loading	
(kg VS/m^3·d)	1.6
Volatile Solids Destroyed (%)	40

Chlorine Contact:

- 30 min HRT
- 8-15 mg/L Chlorine Dosage

FIGURE 5 MODEL EXTENDED AERATION PLANT PROCESS FLOWSHEET

3.2 Costing of Model Plants

The capital and operating costs for each of the model plants shown in Figures 3, 4 and 5 were estimated using the CAPDET model. CAPDET is the acronym for Computer Assisted Procedure for the Design and Evaluation of Wastewater Treatment Systems and was developed by the U.S. Army Corps of Engineers, with support from the U.S. Environmental Protection Agency (Harris et al., 1982).

CAPDET was originally intended as a costing tool for prescreening of wastewater treatment process alternatives on a least-cost basis. It is primarily a planning tool which has been used with considerable success in the United States in facility planning procedures under Section 201 of the Federal Water Pollution Control Act (PL 92-500). McGhee et al. (1983) reported their experiences with CAPDET in the evaluation of a number of wastewater treatment process alternatives for Jefferson Parish, Louisiana. They concluded that, as a screening tool, CAPDET was extremely effective and a useful technique for process selection. However, they cautioned the user against indiscriminate use since the user must still test the program output for reasonableness.

In Canada, use of CAPDET has been less widespread. However, Pineau et al. (1985), after having adapted the CAPDET model to Canadian conditions, determined that construction cost estimates for plants were within $\pm$ 20% of actual construction costs. Operating and maintenance costs were found to be within $\pm$ 30%.

The major strength of the CAPDET model is its capability as a screening and comparison tool to assess the relative costs of various process alternatives. In this investigation, comparative costs rather than actual costs were required to compare the cost-effectiveness of the various phosphorus removal schemes.

3.2.1 CAPDET Simulation. The CAPDET model is based on a FORTRAN program and is made up of numerous sub-routines. Each of the sub-routines contains an algorithm for the design and costing of a particular unit process. Costs are calculated through the program using a unit costing approach.

The CAPDET simulation requires certain input data including unit process specifications, influent waste characteristics, desired effluent quality, and unit cost data.

The CAPDET output includes four components: construction costs of individual unit processes, construction costs of associated facilities, direct and indirect costs, and operation and maintenance (O&M) costs. Included in the direct and indirect costs are overhead and profit (direct costs), and design fees, contingencies, inspection cost, and administration and legal fees (indirect costs).

The total capital cost (TCC) of the project is the sum of the capital cost plus the direct and indirect costs where:

Direct Cost = 0.22 x Capital Cost

Indirect Cost = 0.20 x (Capital Cost + Direct Cost)

The total capital cost of the project is thus,

TCC = 1.464 x Capital Cost

Unit costs can be calculated based on calculated usages of materials, such as concrete for construction, and labour, chemicals and electricity for operation, and user supplied cost information for such items. Unit costs for process equipment such as aerators, mixers, pumps and blowers can be user specified. If cost figures are unavailable, the CAPDET sub-routines will use default values derived for standard size equipment. The default costs in CAPDET were established in 1977. They are updated using appropriate indexes such as the EPA Construction Index, Marshall & Swift Index and the ENR Index.

Use of the CAPDET simulation not only requires considerable process knowledge, but also an understanding of what the simulation does and how calculations are performed. The prediction of construction and O&M costs is only as good as the data input to the model. If local costs are not supplied, a default value is assumed in the calculation. Thus, the more user supplied cost and design information provided as input, the greater will be the accuracy of the cost prediction.

3.2.2 Cost Bases for Model Plants. For the development of the model plant costs, certain assumptions were made with respect to plant hardware, operational procedures and cost factors. These assumptions were provided as input to CAPDET in order to generate capital, O&M and total annual costs for each model plant type at each flow rate.

Operator labour rates were specified at $15.60/h and maintenance labour rates at $18.00/h. Electrical costs were specified at $0.04/kWh. Chlorine costs of $187/tonne were specified.

Sludge disposal costs were based on land disposal of digested sludge at a round trip distance of 32 km. Operating and maintenance costs included loading time and travel time. Capital costs included the disposal vehicle and sludge storage building.

Capital costs were amortized over 20 years at an annual interest rate of 12%. All costs were converted to 1984 Canadian dollars. The cost indices supplied to CAPDET were as follows:

Marshall and Swift Index	=	769.0
ENR Index	=	4175.69
EPA Construction Index (Small City)	=	193.4

3.3 Retrofits to Achieve Phosphorus Control

The relative costs of retrofitting each model plant (three plant types at three design flows) to achieve phosphorus removal by the four EBPR process options (Bardenpho, UCT, A/O and Phostrip) were compared to the costs of achieving the same degree of phosphorus control using chemical precipitation technology. The effluent total phosphorus limits imposed on the processes were 1.0 mg/L and 0.3 mg/L.

3.3.1 Design of Chemical Phosphorus Removal Facilities. To establish a basis for cost comparison, the model primary, conventional activated sludge and extended aeration plants were retrofitted for chemical precipitation of phosphorus.

The retrofit designs of the secondary plants (conventional activated sludge and extended aeration) were based on simultaneous precipitation of phosphorus by the addition of ferric chloride to the aeration tank. Chemical dosages of 8 mg Fe/L to achieve an effluent total phosphorus limit of 1.0 mg/L and 14 mg Fe/L to achieve an effluent total phosphorus limit of 0.3 mg/L were selected. A final effluent granular media filter was included in the flow scheme to achieve the lower total phosphorus limit. In the case of the primary plant configuration, retrofit costs to achieve the lower phosphorus limit (0.3 mg/L) were not developed for the chemical precipitation option (Section 1.2).

In the case of the primary plant, a 40 percent increase in sludge generation was assumed after implementation of chemical phosphorus removal to the 1.0 mg/L discharge limit. A 26 percent increase in sludge generation was assumed for the secondary plants (conventional activated sludge and extended aeration) retrofitted for chemical phosphorus removal to the 1.0 mg/L limit. An overall increase of 42 percent in sludge generation was assumed in the secondary plants retrofitted to achieved the lower (0.3 mg/L) effluent phosphorus limit.

3.3.2 Design of EBPR Processes. The designs of the four EBPR processes were developed based on the design procedures outlined in Section 2. For the totally biological options (Bardenpho, A/O and UCT), the primary clarifiers in the model plants were by-passed to maximize the organic content of the raw wastewater input to the anaerobic basins. The Bardenpho plant was designed as a five-stage process (Figure 2B) for maximum biological nutrient control. The UCT system was designed as a three-stage

system (Figure 2E). A/O was designed as a two-stage system, without biological nitrogen control. In the case of Phostrip, the mainstream biological process did not provide for biological nitrification. These designs were intended to cover the entire range of nutrient control (nitrogen and phosphorus) available with the EBPR processes so that the point at which EBPR processes were cost competitive with chemical phosphorus removal could be identified.

The design of the Bardenpho, A/O and UCT systems to achieve the higher effluent total phosphorus limit (1.0 mg/L) followed two approaches as indicated in Table 12: EBPR plus effluent polishing by dual media filtration, and EBPR plus supplemental ferric chloride addition to reduce the soluble orthophosphorus concentration in the effluent. In the latter case, it was assumed that ferric chloride would be applied only during effluent excursions above the 1.0 mg/L TP limit. The designs of these processes to achieve the lower effluent total phosphorus limit (0.3 mg/L) incorporated both dual media filtration and supplemental ferric chloride addition. The ferric chloride dosage rate was 6 mg Fe/L, equivalent to the difference between the chemical dosages required to achieve 1.0 mg/L TP (8 mg Fe/L) and 0.3 mg/L TP (14 mg Fe/L) in the conventional chemical phosphorus removal facility.

In the case of the Phostrip designs, the plants designed to achieve 1.0 mg/L effluent total phosphorus incorporated only the Phostrip process retrofit. Designs to achieve the lower effluent limit incorporated effluent dual media filtration. The Phostrip process was designed with a recycle stripping tank, with phosphorus precipitation by lime addition in a dedicated reactor-clarifier.

The EBPR process retrofits were designed to maximize the use of existing equipment. For example, in the primary plant retrofit situations, the flow scheme was modified to allow the model plant primary clarifier to be utilized as the secondary clarifier in the retrofitted plant. Existing aeration tankage was reused as much as possible, but compartmentalization was incorporated by the addition of concrete partitions and sluice gates. To provide flexibility in the various reaction zones (anaerobic, anoxic and aerobic), submerged turbine aerators were designed into the retrofitted EBPR processes. Mixing energy in the anoxic and anaerobic reactors was 10 W/m^3. Blower facilities provided the additional oxygen transfer capability required to meet the carbonaceous (and nitrogenous) oxygen demand.

Phostrip plant designs incorporated a bag storage and feeder system for plants with design flows of less than 4.5 ML/d. For larger plants, a bulk lime silo storage system was incorporated. Lime dosage to the stripper overflow, at 240 mg $Ca(OH)_2$/L, and solids

production from the lime precipitation process, at 0.5 kg dry solids per cubic metre of stripper supernatant, were based on stoichiometric calculations.

Based on these design assumptions, retrofit EBPR process flowsheets were developed for the Bardenpho, UCT, A/O and Phostrip systems, using the model primary, conventional activated sludge and extended aeration plant configurations as the starting points. The retrofit process flowsheets are presented in Figures 6 to 17, along with the pertinent design criteria. As the retrofit designs were intended to incorporate existing equipment to the greatest extent possible, the actual process design details (basin hydraulic retention times) varied slightly from case-to-case in order to maximize the use of existing tankage but, at the same time, maintain typical design ranges normally specified for the selected processes. Table 15 presents an index to the EBPR retrofit process design retrofits shown in Figures 6 to 17.

TABLE 15 INDEX OF EBPR RETROFIT PROCESS DESIGNS

EBPR Option	Retrofitted Model Plant		
	Primary	Conventional Activated Sludge	Extended Aeration
Bardenpho	Figure 6	Figure 7	Figure 8
UCT	Figure 9	Figure 10	Figure 11
A/O	Figure 12	Figure 13	Figure 14
Phostrip	Figure 15	Figure 16	Figure 16

3.3.3 Retrofit Process Equipment List. Based on the EBPR process retrofit designs, a list of process equipment and accessories required to complete the retrofit was developed. This equipment list included such hardware as pumps, piping, blowers, mixers, dual media filters, chemical feed equipment, reactor-clarifier equipment and biological reactor tankage. A similar equipment list was developed for the retrofit requirements of the chemical phosphorus removal processes.

3.3.4 Development of Retrofit Costs. Capital and O&M cost curves were developed for each type of equipment included in the retrofit process equipment list for a range of flows up to 40.0 ML/d. These cost curves were based on equipment vendors' information, Canviro's cost data and on cost data included in the CAPDET model. The cost curves for equipment and accessories actually used in the model plant retrofits described in

Sections 3.3.1 and 3.3.2, and cost curves for additional process equipment such as flow equalization facilities and dissolved air flotation thickeners, which were not used in these retrofits but which might be required for a specific application, are appended to this report. An index to the cost curves, indicating which process retrofit used each cost curve, is also provided in the Appendix.

The costs of retrofitting the model plants to achieve phosphorus removal, by either biological or chemical means, were developed by selecting the required retrofit hardware and determining the costs (capital and operating) of this hardware from the pertinent cost curve at the required flow. These retrofit equipment costs were then added to the model plant costs developed from the CAPDET model. Appropriate adjustments to sludge disposal costs were made to reflect changes in sludge production rates. Wherever appropriate, cost credits for existing or redundant equipment were assigned to the retrofit. For example, in situations where the primary clarifier was by-passed, an O&M cost credit was assigned equivalent to the O&M cost associated with the operation of the primary clarifier.

After assignment of costs for all retrofit hardware and allocation of all pertinent cost credits, capital costs, O&M costs and total annual costs were calculated for each retrofit situation. O&M costs and total annual costs were also expressed relative to the mass of phosphorus removed ($ per kg total phosphorus removed). It is emphasized that the costs reported are the additional costs associated with implementing phosphorus removal, and not the total plant costs.

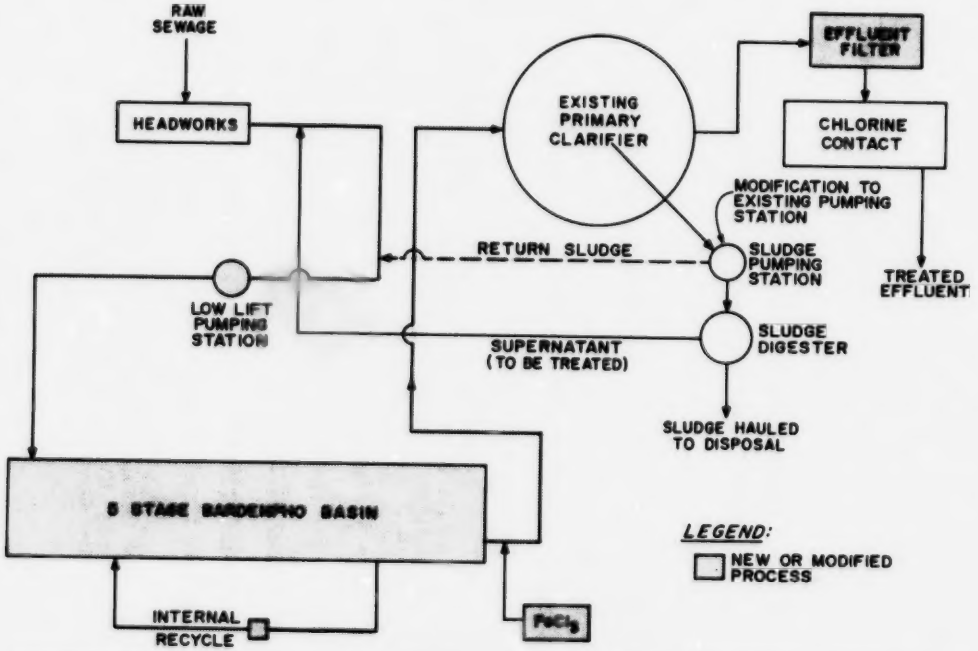
3.3.5 Cost Bases for Retrofit Plants. As much as possible, the cost bases used in the development of the cost curves for phosphorus removal were the same as those used in the development of the model plant costs. The following specific cost assumptions were made for the retrofit situations:

i) Ferric chloride costs were specified as $0.92/kg Fe, including transportation (F.O.B. London, Ontario).
ii) Lime costs were $77/tonne, including transportation (F.O.B. London, Ontario).
iii) No license fees for the EBPR options were included in the retrofit cost calculations.
iv) No cost credits were assigned for nitrogen control in the EBPR plants that achieved nitrogen control concurrent with biological phosphorus removal (Bardenpho and UCT).

Additional details regarding the development of the retrofit accessory cost curves are shown on the cost curves in the Appendix.

MODEL PLANT TYPE : PRIMARY

PROCESS FLOW SHEET



Retrofit Design Criteria

Low Lift Pumping Station

- Peak Flow Capacity

Bardenpho Basin - 5 Stage

Stage 1:	Anaerobic	-	1.0 h HRT
Stage 2:	Anoxic	-	3.5 h HRT
Stage 3:	Aerobic	-	4.5 h HRT
Stage 4:	Anoxic	-	0.5 h HRT
Stage 5:	Aerobic	-	0.5 h HRT
			10.0 h HRT

Internal Recycle Pumping Station

- Stage 3 to Stage 2 @ 400% of Average Flow

Modifications to existing Sludge Pumping Stations

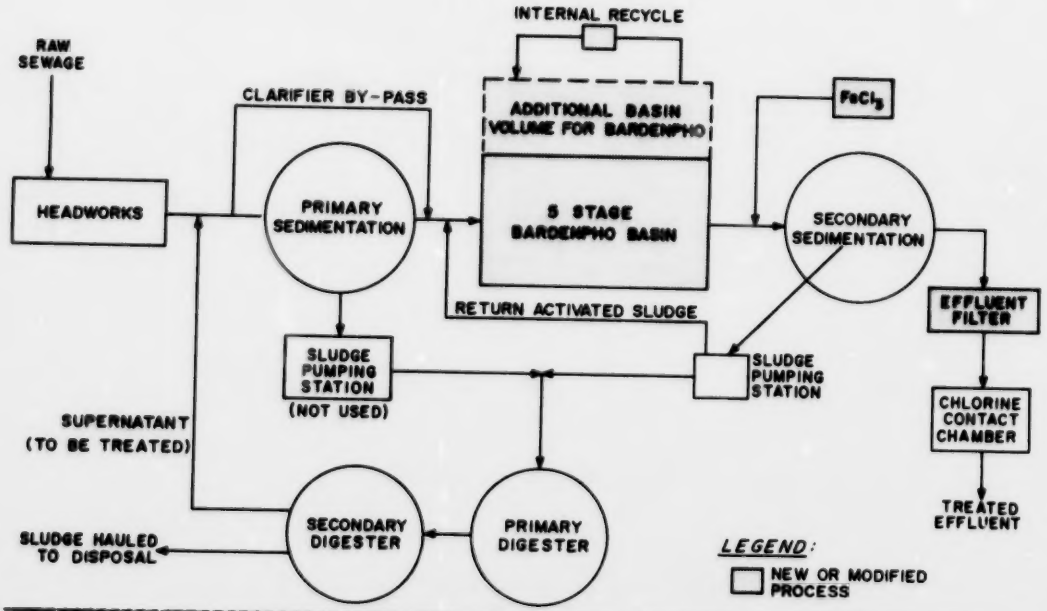
- Modifications as required for sludge return to Bardenpho basin

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 6 RETROFIT OF MODEL PRIMARY PLANT TO BARDENPHO PROCESS

MODEL PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

PROCESS FLOW SHEET



Retrofit Design Criteria

Bardenpho Basin - 5 Stage

Stage 1:	Anaerobic	-	1.0 h HRT
Stage 2:	Anoxic	-	3.4 h HRT
Stage 3:	Aerobic	-	4.5 h HRT
Stage 4:	Anoxic	-	0.5 h HRT
Stage 5:	Aerobic	-	0.5 h HRT
	Total		9.9 h HRT

Stage 2 HRT equals model plant HRT.

Internal Recycle Pumping Station

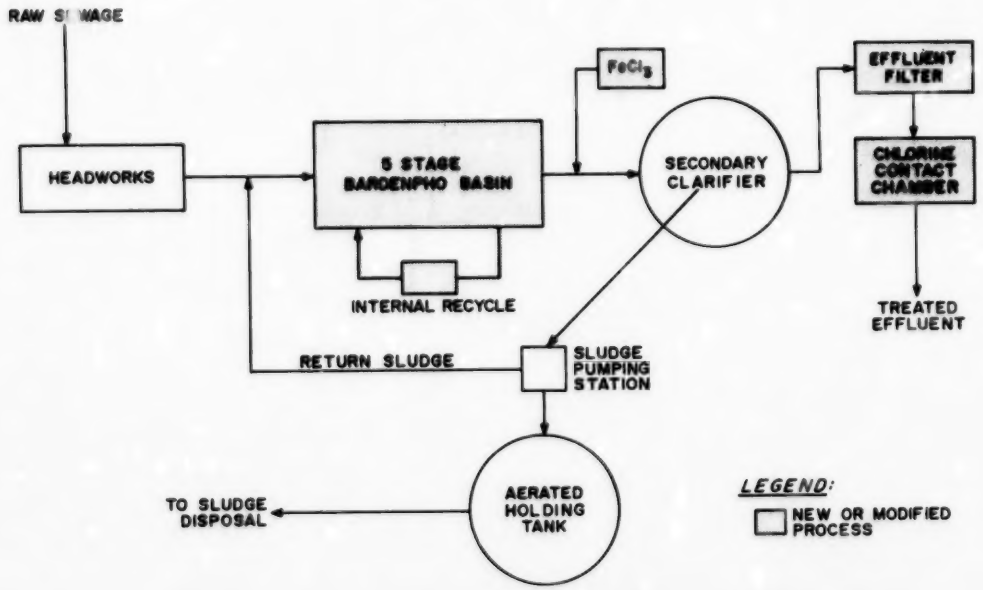
- Stage 3 to Stage 2 @ 400% of Average Flow

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 7 RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO BARDENPHO PROCESS

MODEL PLANT TYPE : EXTENDED AERATION

PROCESS FLOW SHEET



Retrofit Design Criteria

Bardenpho Basin - 5 Stage

Stage 1:	Anaerobic	-	1.25 h HRT
Stage 2:	Anoxic	-	4.33 h HRT
Stage 3:	Aerobic	-	5.6 h HRT
Stage 4:	Anoxic	-	0.6 h HRT
Stage 5:	Aerobic	-	0.6 h HRT
	Total		12.4 h HRT

12.4 h HRT equals model plant HRT.

Internal Recycle Pumping Station

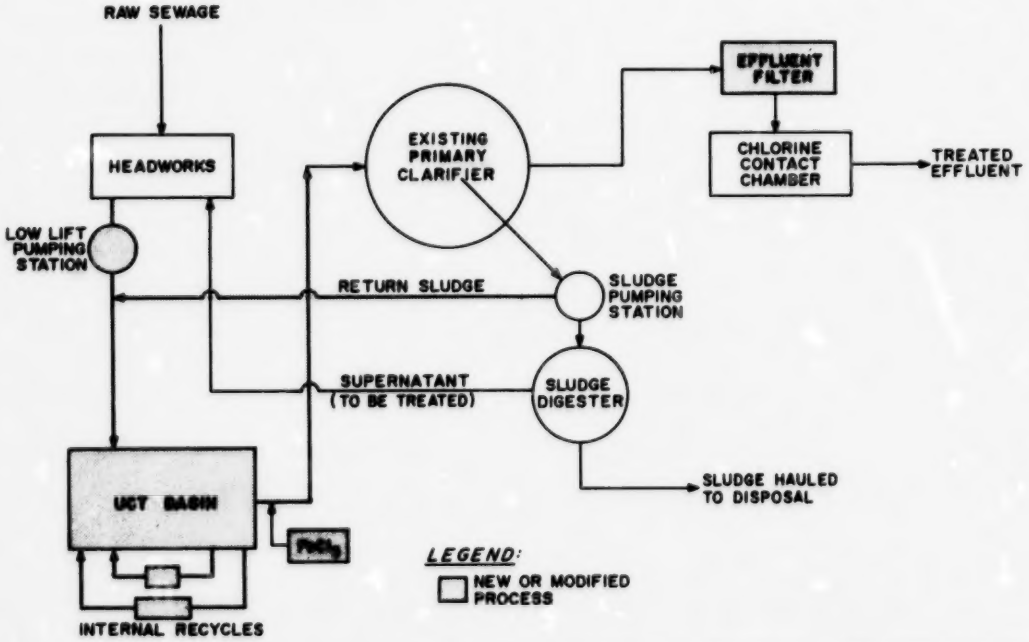
- Stage 3 to Stage 2 @ 400% of Average Flow

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 8 RETROFIT OF MODEL EXTENDED AERATION PLANT TO BARDENPHO PROCESS

MODEL PLANT TYPE : PRIMARY

PROCESS FLOW SHEET



Retrofit Design Criteria

Low Lift Pumping Station

- Peak Flow Capacity

UCT Basin

Stage 1:	Anaerobic	-	1.5 h HRT
Stage 2:	Anoxic	-	2.5 h HRT
Stage 3:	Aerobic	-	6.0 h HRT
	Total		10.0 h HRT

Internal Recycle Pumping Stations

- Designed @ 400% of Average Flow

Existing Sludge Pumping Station Modifications

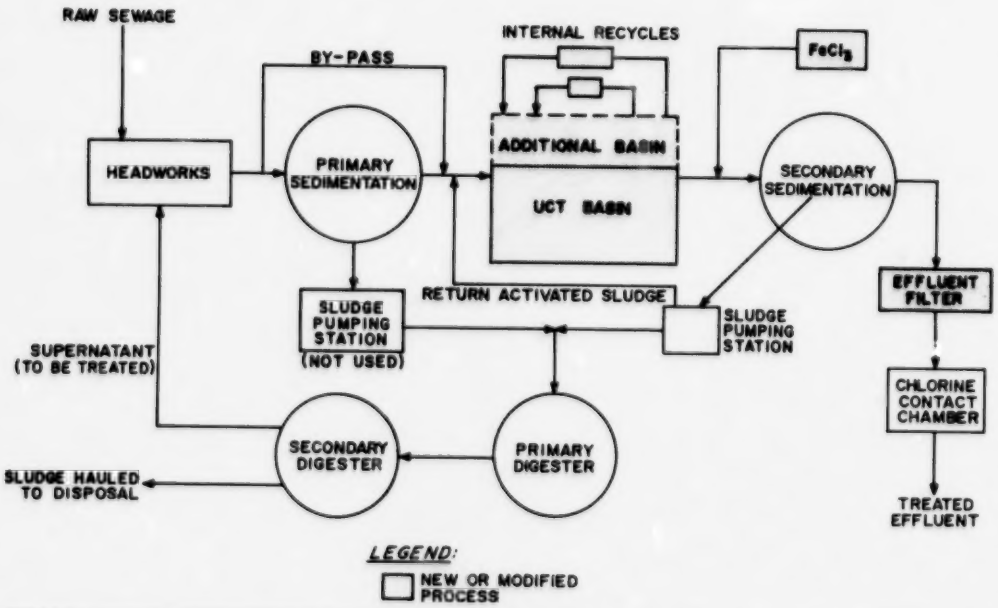
- Modified as required to return sludge to UCT basin.

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 9 RETROFIT OF MODEL PRIMARY PLANT TO UCT PROCESS

MODEL PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

PROCESS FLOW SHEET



Retrofit Design Criteria

UCT Basin

Stage 1:	Anaerobic	-	1.5 h HRT
Stage 2:	Anoxic	-	3.4 h HRT
Stage 3:	Aerobic	-	6.0 h HRT
	Total		10.9 h HRT

Stage 2 HRT equals model plant HRT.

Internal Recycle Pumping Stations

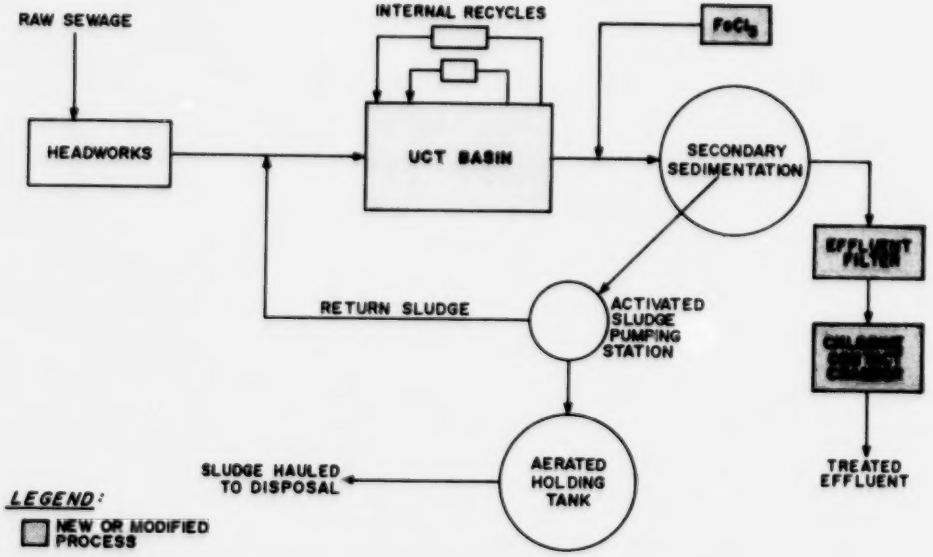
- Designed @ 400% of Average Flow

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 10 RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE TO UCT PROCESS

MODEL PLANT TYPE : EXTENDED AERATION

PROCESS FLOW SHEET



Retrofit Design Criteria

UCT Basin

Stage 1:	Anaerobic	-	2.33 h HRT
Stage 2:	Anoxic	-	3.88 h HRT
Stage 3:	Aerobic	-	6.2 h HRT
	Total		12.4 h HRT

12.4 h HRT equals HRT value of model plant.

Internal Recycle Pumping Stations

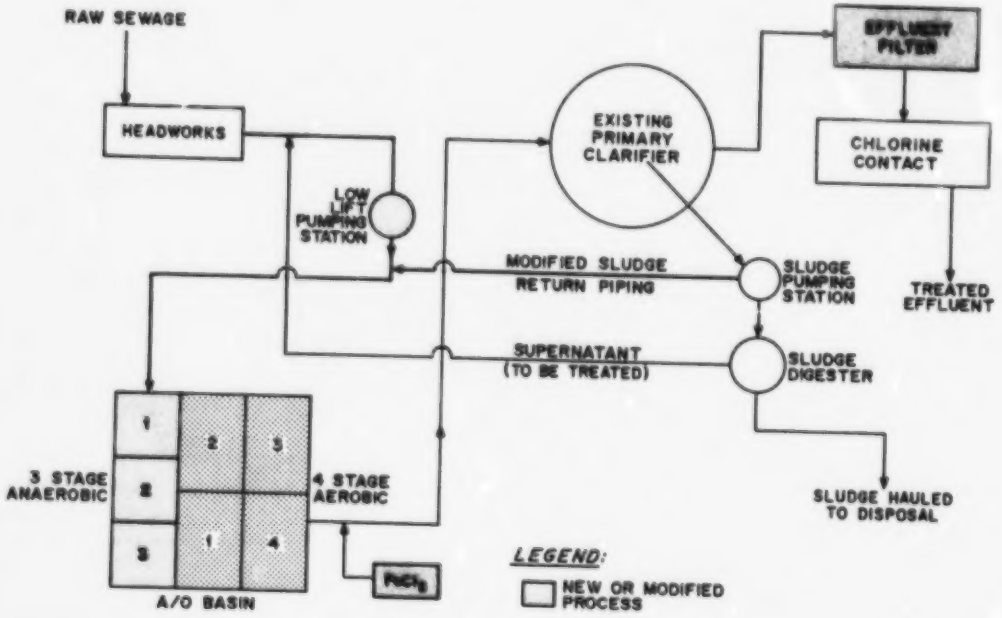
- Designed @ 400% of Average Flow

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 11 RETROFIT OF MODEL EXTENDED AERATION PLANT TO UCT PROCESS

MODEL PLANT TYPE : PRIMARY

PROCESS FLOW SHEET



Retrofit Design Criteria

Low Lift Pumping Station

- Peak Flow Capacity

A/O Basin

- 3 Stage Anaerobic HRT - 1.2 h HRT
- 4 Stage Aerobic HRT - 2.2 h HRT

Existing Sludge Pumping Station

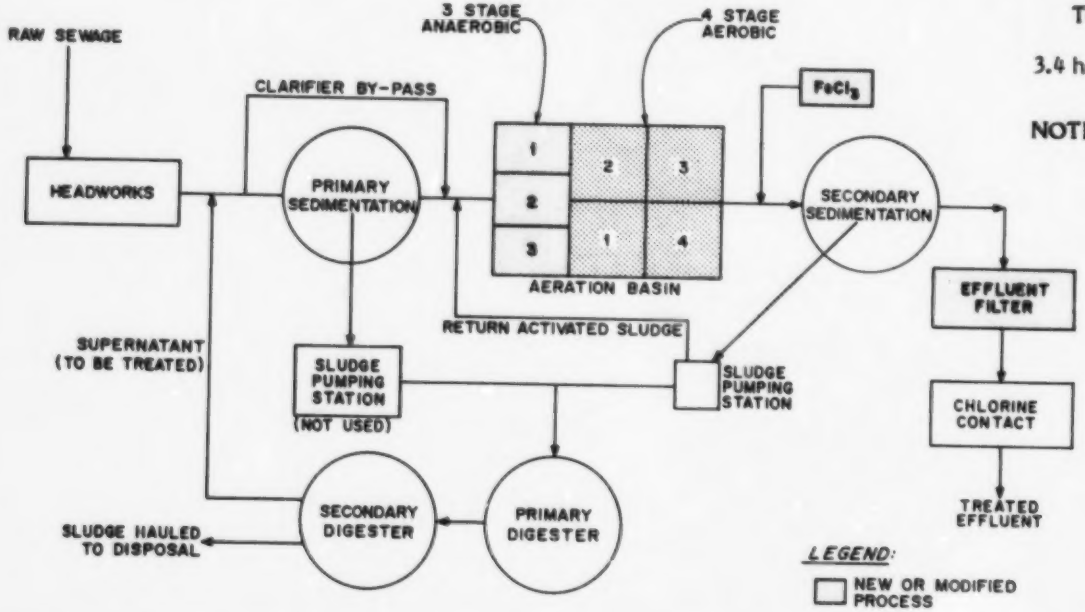
- Modification to pumping station as required for additional sludge return to A/O basin.

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 12 RETROFIT OF MODEL PRIMARY PLANT TO A/O PROCESS

MODEL PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

PROCESS FLOW SHEET



Retrofit Design Criteria

A/O Basin

- 3 Stage Anaerobic HRT	- 1.2 h HRT
4 Stage Aerobic HRT	- 2.2 h HRT
Total	3.4 h HRT

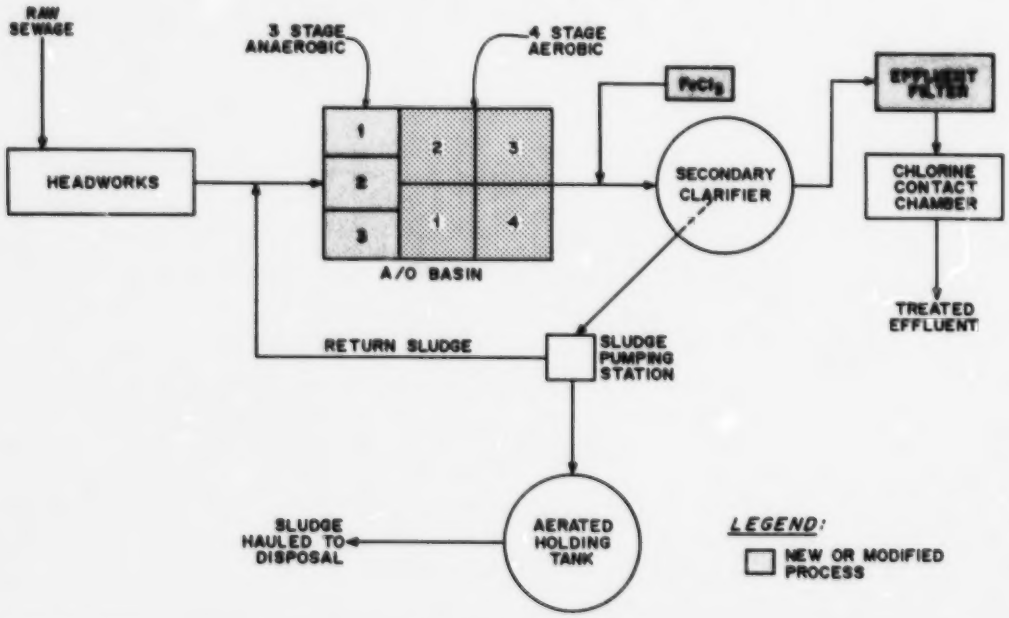
3.4 h HRT equals HRT of model plant.

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 13 RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO A/O PROCESS

MODEL PLANT TYPE : EXTENDED AERATION

PROCESS FLOW SHEET



Retrofit Design Criteria

A/O Basin

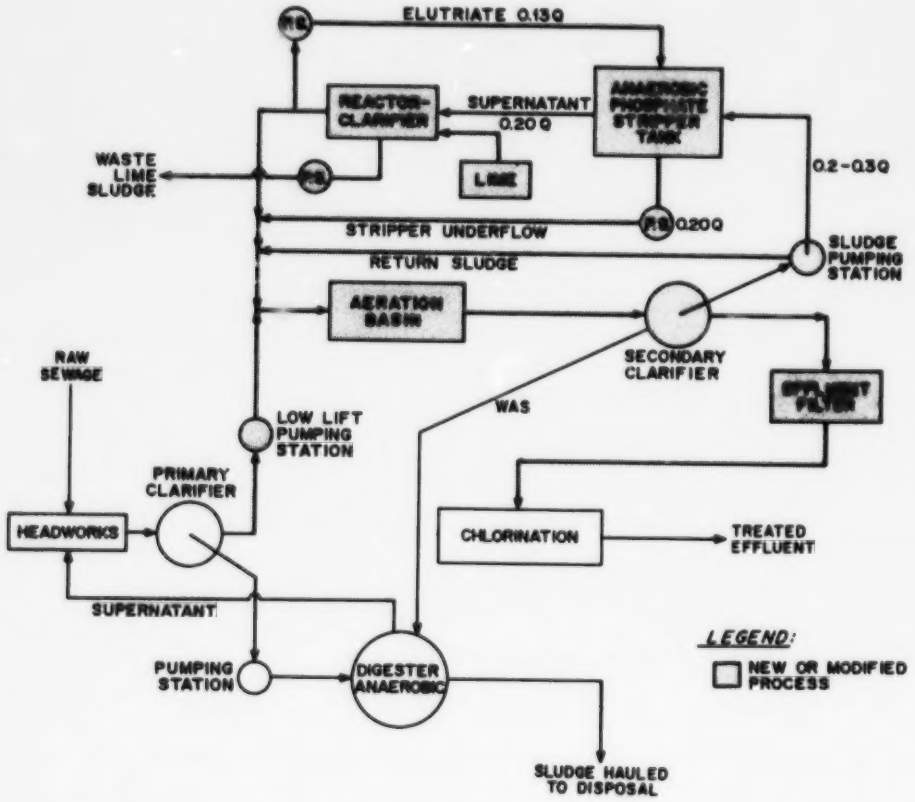
- 3 Stage Anaerobic HRT - 1.2 h HRT
- 4 Stage Aerobic HRT - 2.2 h HRT

NOTE: $FeCl_3$ or filter required to achieve 1.0 mg/L TP. $FeCl_3$ and filter required to achieve 0.3 mg/L TP.

FIGURE 14 RETROFIT OF MODEL EXTENDED AERATION PLANT TO A/O PROCESS

MODEL PLANT TYPE : PRIMARY

PROCESS FLOW SHEET



Retrofit Design Criteria

Low Lift Pumping Station

- Peak Flow Capacity

Aeration Basin

- Organic Loading Rate ($g\ BOD/m^3 \cdot h$) = 13-30
- F/M (d^{-1}) = 0.2-0.5

Secondary Clarifier

- Surface Overflow Rate @ Peak Flow ($m^3/m^2 \cdot d$) = 35
- Sidewater Depth (m) = 3.6-4.6

Activated Sludge Pumping Station

- Average Flow Capacity

Anaerobic Phosphate Stripper Tank

- 10 h HRT
- Inflow @ 20-30% Average Flow

Reactor-Clarifier

- Surface Overflow Rate @ Average Flow ($m^3/m^2 \cdot d$) = 49
- Underflow Pumping Capacity @ 10-20% Average Flow
- Elutriate Pumping Capacity @ 10-30% Average Flow

Chemical Lime Feed System

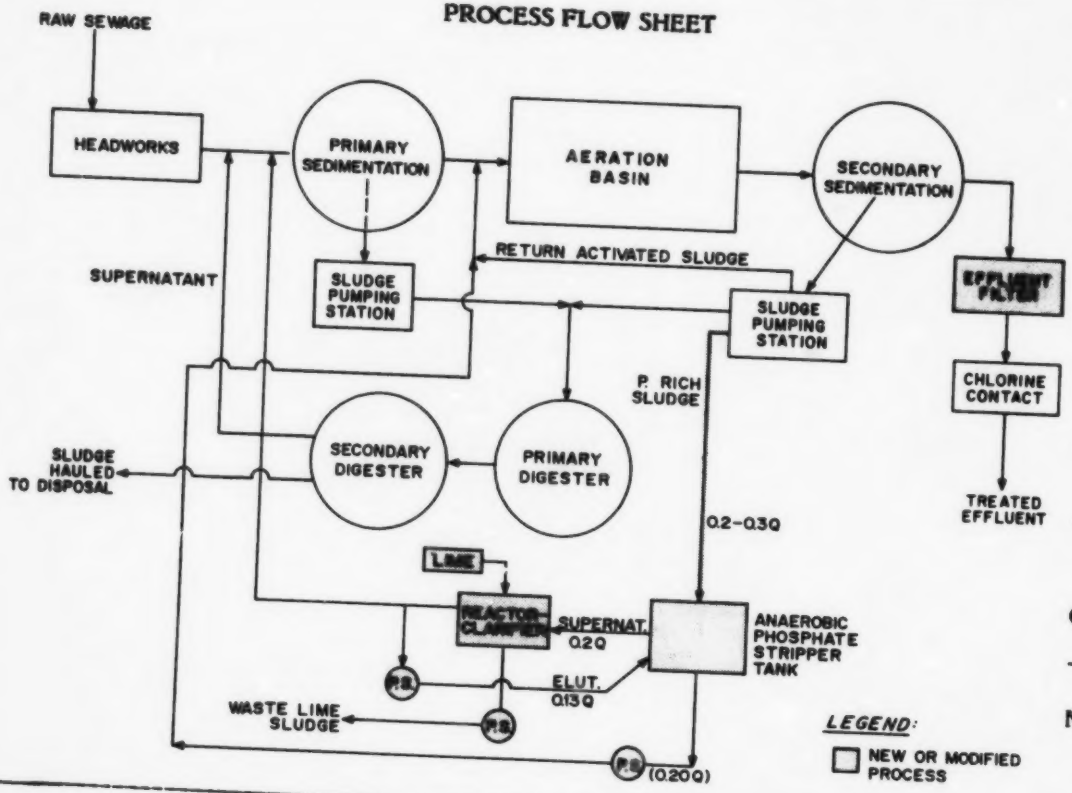
- 240 mg/L Hydrated Lime

NOTE: Effluent filter only required to achieve 0.3 mg/L TP.

FIGURE 15 RETROFIT OF MODEL PRIMARY PLANT TO PHOSTRIP PROCESS

MODEL PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

PROCESS FLOW SHEET



Retrofit Design Criteria

Existing Activated Sludge Pumping Station Modifications

- Modifications as required to transfer sludge to anaerobic phosphate stripper tank.

Anaerobic Phosphate Stripper Tank

- 10 h HRT
- Inflow @ 20-30% Average Flow

Reactor-Clarifier

- Surface Overflow Rate @ Average Flow ($m^3/m^2.d$) = 49
- Underflow Pumping Capacity @ 10-20% Average Flow
- Elutriate Pumping Capacity @ 10-30% Average Flow

Chemical Feed System

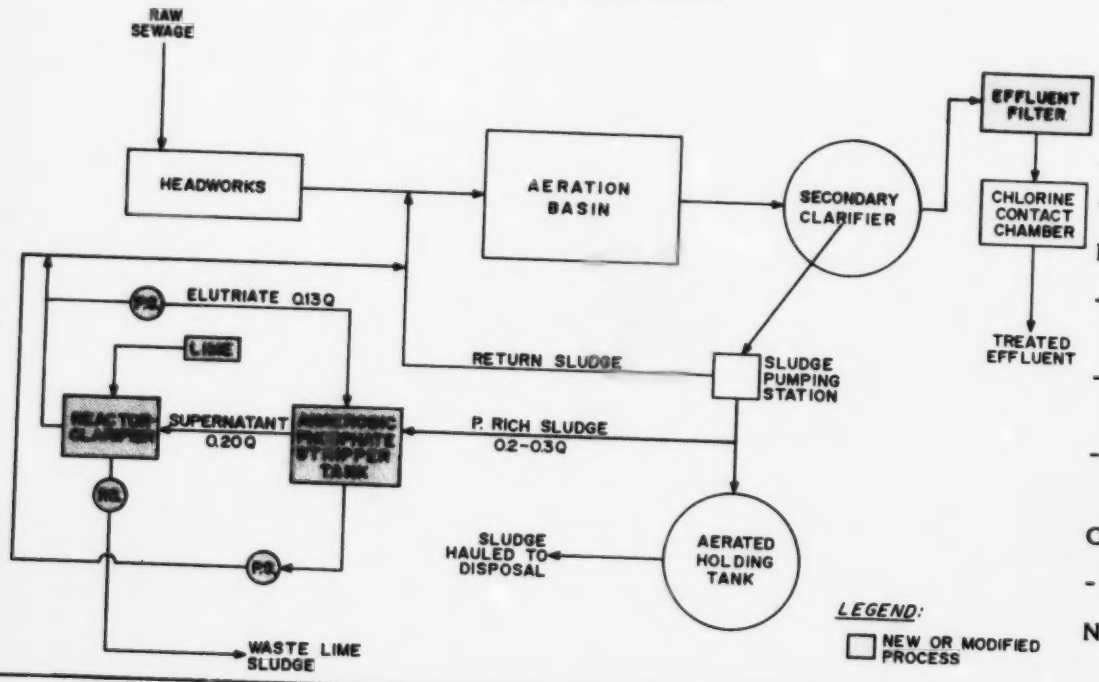
- 240 mg/L Hydrated Lime

NOTE: Effluent filter only required to achieve 0.3 mg/L TP.

FIGURE 16 RETROFIT OF MODEL CONVENTIONAL ACTIVATED SLUDGE PLANT TO PHOSTRIP PROCESS

MODEL PLANT TYPE : EXTENDED AERATION

PROCESS FLOW SHEET



Retrofit Design Criteria

Existing Activated Sludge Pumping Station Modifications

- Modifications as required to transfer sludge to anaerobic phosphate stripper tank.

Anaerobic Phosphate Stripper Tank

- 10 h HRT
- Inflow @ 20-30% Average Flow

Reactor-Clarifier

- Surface Overflow Rate @ Average Flow ($m^3/m^2 \cdot d$) = 49
- Underflow Pumping Capacity @ 10-20% Average Flow
- Elutriate Pumping Capacity @ 10-30% Average Flow

Chemical Feed System

- 240 mg/L Hydrated Lime

NOTE: Effluent filter only required to achieve 0.3 mg/L TP.

FIGURE 17 RETROFIT OF MODEL EXTENDED AERATION PLANT TO PHOSTRIP PROCESS

3.4 New Plant Costs

In addition to developing retrofit cost comparisons for EBPR process options and chemical precipitation options in existing plants, cost comparisons were developed for new facilities. The five new plant design configurations used were:

i) conventional activated sludge with simultaneous precipitation of phosphorus using ferric chloride;
ii) Bardenpho process;
iii) UCT process;
iv) A/O process; and
v) Phostrip process in conjunction with non-nitrifying conventional activated sludge.

The same total phosphorus effluent concentration limits (1.0 mg/L and 0.3 mg/L) were imposed on the new facilities as had been applied to the retrofit facilities. The cost comparisons used design flows of 4.5 ML/d, 13.6 ML/d and 36.4 ML/d, the same design capacities used in the conventional activated sludge retrofit examples.

All of the fundamental design and cost assumptions applied in the retrofit situations were applied in the analysis of the new plant costs. CAPDET and the cost curves presented in the Appendix were the principal sources of cost data. Capital, O&M and total annual costs were developed for each facility design. The costs represent the total associated with construction and operation of the facility, not the costs associated with the operation of the phosphorus removal component of the facility, as was the case for the retrofit facilities. O&M and total annual costs were also calculated on a $ per kg phosphorus removed basis.

As one component of the new plant cost analysis, the sensitivity of the costs of the simultaneous chemical precipitation technology to chemical dosage and chemical cost was evaluated. The chemical dosage to achieve the 1.0 mg/L effluent limit was increased by 100 percent to 16 mg/L Fe and the effect on annual O&M and total annual costs was evaluated. The impact of a 100 percent increase in ferric chloride cost (to $1.84 per kg Fe) on the O&M and total annual costs was also evaluated. In addition, the effect of using alum in the simultaneous precipitation process was assessed. In this analysis, an alum cost of $220 per tonne (dry alum basis) was assumed, including transportation. This cost is appropriate for either Western Canada (Calgary) or Southwestern Ontario (London). Alum dosages of 75 mg/L and 125 mg/L as alum were assumed to achieve effluent total phosphorus limits of 1.0 mg/L and 0.3 mg/L, respectively.

4 ECONOMIC ASSESSMENT OF PHOSPHORUS REMOVAL TECHNOLOGIES

The retrofit and new facility costs for each of the EBPR processes (Bardenpho, UCT, A/O and Phostrip) and for the simultaneous chemical precipitation technology, developed as outlined in Section 3, are presented in this section.

4.1 Retrofit Costs

As noted previously, the retrofit costs represent the additional costs associated with implementing phosphorus removal to two effluent total phosphorus limits (1.0 mg/L and 0.3 mg/L) in each of the model plants described in Section 3. In achieving the upper phosphorus limit, two scenarios were considered for the totally biological process options (Bardenpho, UCT, A/O):

i) EBPR in conjunction with effluent filtration; and

ii) EBPR in conjunction with supplemental chemical addition as required to achieve 1.0 mg/L total phosphorus in the final effluent.

Incremental capital costs, incremental O&M costs ($/annum) and incremental annual costs ($/annum) are presented for each retrofit situation. In addition, the incremental O&M costs and annnual costs are expressed on a "$ per kg total phosphorus removed" basis.

4.1.1 Chemical Precipitation Process. Costs for retrofitting the model primary plants for phosphorus removal by chemical precipitation to the 1.0 mg/L total phosphorus limit are presented in Table 16. Costs of achieving the lower effluent total phosphorus limit (0.3 mg/L) in a chemical precipitation retrofit of the model primary plant were not developed because this approach was not considered to be technically feasible.

Costs for retrofitting the model conventional activated sludge plants for simultaneous chemical precipitation are presented in Table 17. The lower phosphorus limit (0.3 mg/L) was achieved by increased chemical dosage in combination with effluent filtration. Effluent filtration resulted in a substantial increase ($900 000 to $2 898 600) in the capital cost of the plant retrofits. Amortization of this capital cost was responsible for a substantial fraction of the increase in total annual costs compared to the cost of achieving the higher effluent total phosphorus limit.

Costs of retrofitting the model extended aeration plants for simultaneous chemical precipitation are presented in Table 18. The observations made with respect to the impact of effluent filtration on the costs of achieving the lower phosphorus limit in conventional activated sludge plants also apply to the extended aeration plant retrofits.

TABLE 16 COSTS TO RETROFIT PRIMARY PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY 2.3 ML/d; EFFLUENT TOTAL PHOSPHORUS LIMIT 1.0 mg/L	HYDRAULIC PLANT CAPACITY 9.1 ML/d; EFFLUENT TOTAL PHOSPHORUS LIMIT 1.0 mg/L	HYDRAULIC PLANT CAPACITY 36.4 ML/d; EFFLUENT TOTAL PHOSPHORUS LIMIT 1.0 mg/L
Capital	$	126 000	132 000	159 600
O&M	$/annum	22 900	76 500	294 300
	$/kg TP_R	6.91	5.76	5.54
Total Annual	$/annum	39 800	94 100	315 600
	$/kg TP_R	12.01	7.09	5.95

TP_R = total phosphorus removed.

4.1.2 Bardenpho Process. Costs for retrofitting the model primary plants to achieve biological nutrient control using the Bardenpho process are summarized in Table 19. Not surprisingly, the provision of supplemental chemical addition on an "as-needed basis" is considerably more cost-effective than the addition of granular media filtration for effluent polishing.

Costs for retrofitting the model conventional activated sludge plants to achieve biological nutrient control using the Bardenpho process are summarized in Table 20. Cost data for retrofitting the model extended aeration plants with the Bardenpho process are summarized in Table 21.

Retrofitting primary plants to achieve phosphorus removal using the Bardenpho process is more capital cost intensive than retrofitting conventional activated sludge plants of the same hydraulic capacity and to the same effluent total phosphorus limit. The high capital cost of the primary plant retrofits is associated with the biological treatment components (aeration tanks and clarification equipment) that must be supplied.

The economy of scale associated with the secondary plant retrofits is evident from the cost data in Tables 20 and 21. The O&M and total annual costs, calculated on the basis of kilograms phosphorus removed, decrease significantly as plant capacity increases.

TABLE 17 COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL

		HYDRAULIC PLANT CAPACITY					
		4.5 ML/d		13.6 ML/d		36.4 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	129 600	1 012 400	136 200	1 752 900	159 600	3 058 200
O&M	$/annum	37 400	82 800	103 900	214 600	253 400	493 300
	$/kg TP_R	5.64	10.63	5.22	9.18	4.77	7.90
Total Annual	$/annum	54 800	218 300	122 100	449 200	274 800	902 700
	%/kg TP_R	8.27	28.03	6.14	19.22	5.17	14.46

TABLE 18 COSTS TO RETROFIT EXTENDED AERATION PLANTS FOR CHEMICAL PHOSPHORUS REMOVAL

		HYDRAULIC PLANT CAPACITY					
		0.91 ML/d		1.6 ML/d		3.2 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	124 000	514 000	125 000	600 000	125 000	745 000
O&M	$/annum	7 700	20 400	12 300	30 700	21 600	34 500
	$/kg TP_R	5.79	13.07	5.26	11.18	4.62	6.28
Total Annual	$/annum	24 300	89 200	29 000	111 000	38 300	134 200
	%/kg TP_R	18.28	57.14	12.41	40.44	8.20	24.44

TP_R = total phosphorus removed.

TABLE 19 COSTS TO RETROFIT PRIMARY PLANTS WITH BARDENPHO TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		2.3 ML/d			9.1 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	1 608 000	1 117 800	1 759 800	3 124 600	1 999 060	3 283 000	7 706 400	4 993 900	7 987 900
O&M	$/annum	101 500	85 000	113 000	194 000	163 000	239 700	433 000	361 000	617 600
	$/kg TP_R	30.63	25.62	29.02	14.62	12.28	15.37	8.16	6.80	9.90
Total Annual	$/annum	316 900	234 651	348 600	612 300	430 620	679 200	1 464 700	1 029 600	1 675 000
	$/kg TP_R	95.61	70.71	89.52	46.14	32.44	43.56	27.59	19.39	26.85

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 20 COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH BARDENPHO TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		4.5 ML/d			13.6 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	1 813 200	1 085 500	1 968 700	3 246 000	1 789 400	3 409 400	5 766 000	3 053 500	5 957 500
O&M	$/annum	67 600	46 100	91 500	124 500	84 500	193 200	209 100	137 100	393 700
	$/kg TP_R	10.20	6.95	11.75	6.26	4.24	8.27	3.93	2.58	6.30
Total Annual	$/annum	310 400	191 400	355 100	539 100	324 100	649 630	981 100	545 900	1 191 300
	$/kg TP_R	46.83	28.84	45.60	28.12	16.28	27.80	18.46	10.28	19.08

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 21 COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH BARDENPHO TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		0.91 ML/d			1.6 ML/d			3.2 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	918 000	598 800	1 066 800	1 140 000	720 000	1 290 000	1 446 000	852 000	1 596 000
O&M	$/annum	33 100	26 100	38 300	43 700	32 700	51 100	48 300	31 300	59 200
	$/kg TP_R	24.90	19.67	24.54	18.71	14.07	18.62	10.34	6.74	10.78
Total Annual	$/annum	156 000	106 300	181 100	196 300	129 100	223 800	241 900	145 400	272 900
	$/kg TP_R	117.38	80.07	116.03	84.04	55.57	81.53	51.77	31.29	49.70

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

The costs of retrofitting the model plants to achieve phosphorus control by the Bardenpho process are compared to the costs of achieving similar control by chemical precipitation technology in Figure 18 (effluent total phosphorus limit 1.0 mg/L) and Figure 19 (effluent total phosphorus limit 0.3 mg/L). These cost comparisons are based on the total annual costs (per kg total phosphorus removed) for the retrofit situations.

The cost data in Figure 18 demonstrate the cost advantage associated with achieving the 1.0 mg/L effluent total phosphorus limit by chemical supplementation compared to the costs of achieving this limit by effluent filtration. However, in all cases, chemical precipitation technology is more cost-effective than the retrofit of the Bardenpho process on the basis of total annual costs. In the largest conventional activated sludge plant evaluated (36.4 ML/d), the least expensive Bardenpho retrofit is approximately twice as costly as chemical precipitation. The annual O&M costs of the Bardenpho retrofit in this situation are only half of the annual O&M costs of chemical precipitation. The higher total annual costs of the Bardenpho process retrofits are related to the high capital costs associated with the biological reactors required for the five-stage process in order to achieve complete nitrogen control. No nitrogen control is provided by the chemical precipitation technology.

A similar situation exists in retrofitting the model plants to achieve the lower phosphorus limit (0.3 mg/L), as shown in Figure 19. Chemical precipitation is the more cost-effective alternative in all cases, based on total annual costs. Bardenpho technology is approximately 32 percent more expensive in the conventional activated sludge retrofit at the largest scale evaluated. However, annual O&M costs of Bardenpho technology are 20 percent lower than annual O&M costs of chemical precipitation in this case.

As noted in Section 3, the cost data presented do not include any license fees for Bardenpho technology. It is our understanding that, at the present time, no such fees would be applicable to the Bardenpho process in Canada.

4.1.3 UCT Process. Costs for retrofitting the model primary plants to achieve biological nutrient control using the UCT process are summarized in Table 22. Similar cost summaries for the model conventional activated sludge plants and extended aeration plants are presented in Table 23 and Table 24, respectively.

The costs associated with achieving the upper effluent total phosphorus limit (1.0 mg/L) by effluent filtration and by supplemental chemical addition are presented. As was the case in the cost analysis for the Bardenpho process, supplemental chemical addition to achieve the effluent quality limit is considerably more cost-effective, as the high capital cost associated with the effluent filter is not incurred.

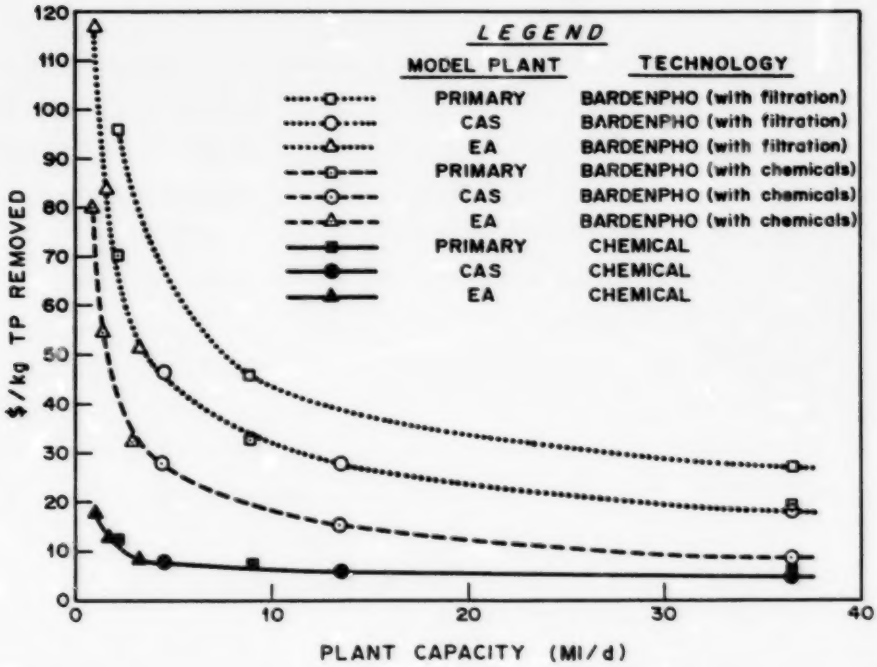


FIGURE 18 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR BARDENPHO AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS

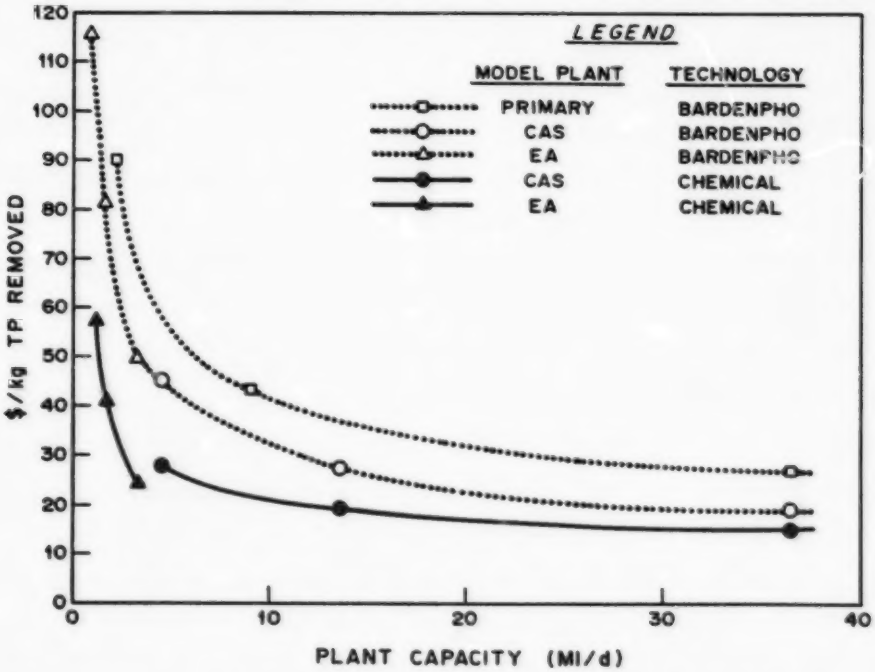


FIGURE 19 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR BARDENPHO AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS

TABLE 22 COSTS TO RETROFIT PRIMARY PLANTS WITH UCT TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY: 2.3 ML/d, EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L) 1.0*	1.0**	0.3	9.1 ML/d 1.0*	1.0**	0.3	36.4 ML/d 1.0*	1.0**	0.3
Capital	$	1 568 800	1 078 000	1 720 000	2 943 400	1 817 800	3 101 800	7 066 800	4 354 300	7 258 300
O&M	$/annum	112 000	95 500	123 500	206 000	175 000	251 700	437 800	365 800	622 400
	$/kg TP_R	33.80	28.78	31.72	15.52	13.18	16.14	8.25	6.89	9.98
Total Annual	$/annum	322 000	239 800	353 800	600 100	418 400	667 000	1 383 900	948 800	1 594 100
	$/kg TP_R	97.16	72.27	90.86	45.22	31.52	42.77	26.07	17.87	25.56

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 23 COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH UCT TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY: 4.5 ML/d, EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L) 1.0*	1.0**	0.3	13.6 ML/d 1.0*	1.0**	0.3	36.4 ML/d 1.0*	1.0**	0.3
Capital	$	1 875 600	1 147 900	2 031 100	3 186 000	1 729 400	3 349 400	5 670 000	2 957 500	5 861 500
O&M	$/annum	79 200	57 700	103 100	138 500	98 500	207 200	214 900	142 900	397 600
	$/kg TP_R	11.95	8.69	13.24	6.97	4.95	8.87	4.04	2.69	6.37
Total Annual	$/annum	330 300	211 400	375 000	565 000	330 000	655 600	974 000	538 853	1 182 300
	$/kg TP_R	49.83	31.85	48.15	28.41	16.58	28.06	18.33	10.15	18.93

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 24 COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH UCT TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY: 0.91 ML/d, EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L) 1.0*	1.0**	0.3	1.6 ML/d 1.0*	1.0**	0.3	3.2 ML/d 1.0*	1.0**	0.3
Capital	$	954 000	634 800	1 102 800	1 152 000	732 000	1 302 000	1 512 000	918 000	1 662 000
O&M	$/annum	43 100	36 100	48 300	52 700	41 700	60 100	60 300	43 290	71 200
	$/kg TP_R	32.43	27.20	30.94	22.56	17.94	21.89	12.91	9.32	12.97
Total Annual	$/annum	170 800	121 100	195 900	206 900	139 700	234 400	262 700	166 200	293 700
	$/kg TP_R	128.52	91.23	125.30	88.57	60.13	85.39	56.23	35.77	53.50

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

The capital costs of the UCT process retrofits and the Bardenpho process retrofits are comparable in most situations. The cost savings associated with the reduced biological reactor requirements of the three-stage UCT process compared to the five-stage Bardenpho process appear to be negated by the costs associated with the additional internal recycle requirements of the UCT process. The annual O&M costs for the UCT process retrofits are generally higher than comparable costs for the Bardenpho process retrofits, because of the additional O&M costs associated with the additional recycle pumping.

The total annual costs for retrofitting the model plants to the UCT process are compared to the costs of retrofitting chemical precipitation technology in Figure 20 (effluent total phosphorus limit 1.0 mg/L) and Figure 21 (effluent total phosphorus limit 0.3 mg/L). As was the case for the retrofit of Bardenpho technology, chemical precipitation technology is more cost-effective than UCT technology in all situations evaluated on the basis of retrofit total annual costs. Achieving the upper total phosphorus limit by retrofitting a conventional activated sludge plant of 36.4 ML/d capacity with the UCT process is at least twice as expensive as retrofitting the same plant with chemical precipitation technology. At the lower phosphorus limit, the cost advantage of the chemical precipitation process is reduced.

Based on the annual O&M costs of the retrofit, the UCT process becomes less expensive when the hydraulic capacity of the conventional activated sludge plant is between 4.5 ML/d and 13.6 ML/d and the effluent total phosphorus limit is 0.3 mg/L or if supplemental chemical addition is utilized instead of filtration to achieve an effluent total phosphorus limit of 1.0 mg/L.

The UCT process has not been proven in North America at full scale. No license fees would be applicable to its use in Canada. The UCT process provides for biological control of both nitrogen and phosphorus, whereas the comparative chemical precipitation process would provide only phosphorus control.

4.1.4 A/O Process. Cost summaries for model plant retrofits to the A/O process are presented in Table 25, Table 26 and Table 27 for the model primary, conventional activated sludge and extended aeration plants, respectively. Costs for achieving the 1.0 mg/L effluent total phosphorus limit by effluent filtration and by supplemental chemical addition are included. As was the case with the other two totally biological EBPR processes (Bardenpho and UCT), supplemental chemical addition creates considerable cost savings compared to the use of effluent filtration.

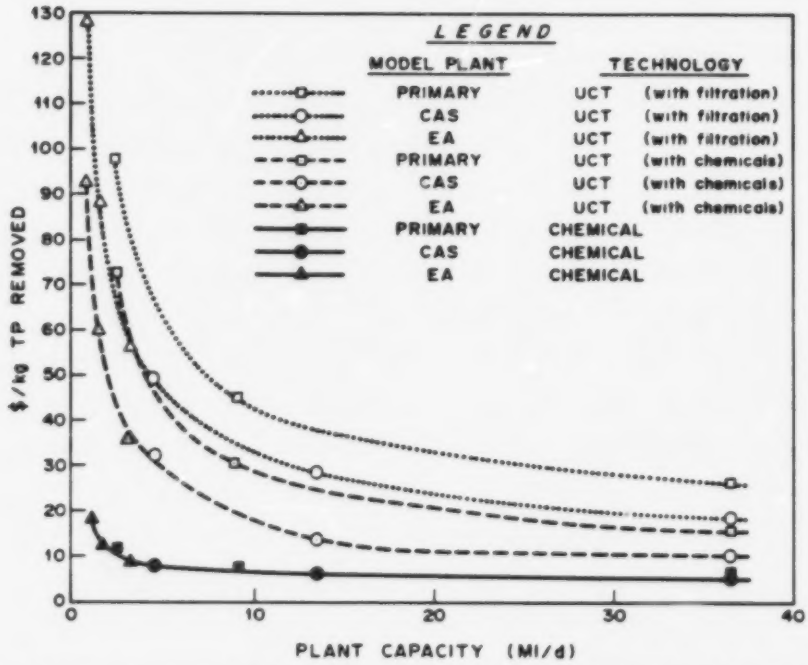


FIGURE 20 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR UCT AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS

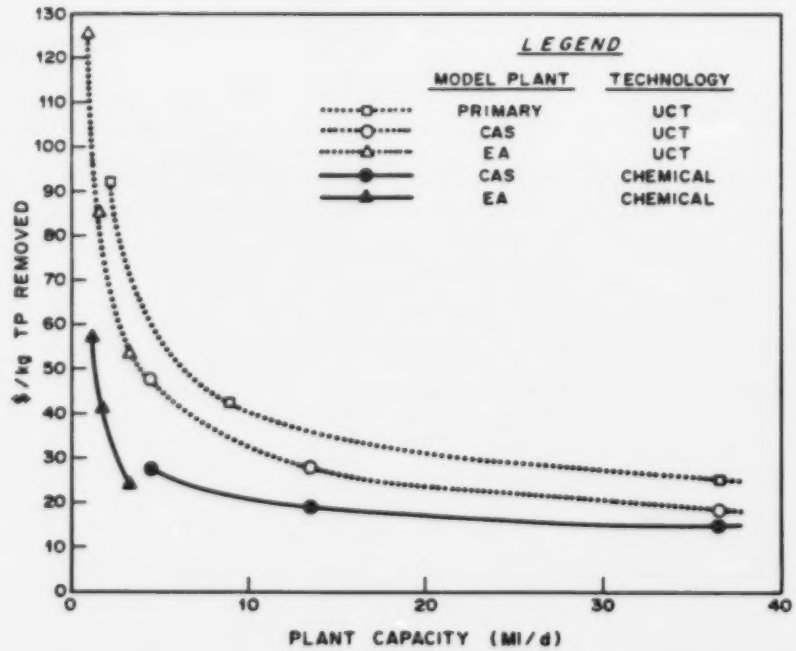


FIGURE 21 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR UCT AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS

TABLE 25 COSTS TO RETROFIT PRIMARY PLANTS WITH A/O TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		2.3 ML/d			9.1 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	1 140 600	649 800	1 291 800	2 227 000	1 101 400	2 385 400	4 759 200	2 046 700	4 930 700
O&M	$/annum	75 500	59 000	87 000	144 000	113 000	189 700	338 000	266 000	522 600
	$/kg TP_R	22.78	17.78	22.34	10.85	8.51	12.16	6.37	5.01	8.38
Total Annual	$/annum	228 200	146 000	259 900	442 200	260 450	509 100	975 200	540 000	1 185 400
	$/kg TP_R	68.86	44.00	66.74	33.32	19.62	32.65	18.37	10.17	19.00

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 26 COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH A/O TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		4.5 ML/d			13.6 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	1 272 600	544 900	1 428 100	2 252 400	795 800	2 415 800	3 828 000	1 115 500	4 109 500
O&M	$/annum	37 100	15 600	61 000	66 500	26 500	133 200	115 100	43 100	297 800
	$/kg TP_R	5.60	2.35	7.83	3.34	1.33	5.79	2.17	0.81	4.77
Total Annual	$/annum	207 500	88 600	252 200	368 100	133 000	458 600	627 600	192 400	835 900
	$/kg TP_R	31.31	13.34	32.38	18.51	6.68	19.63	11.81	3.62	13.39

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 27 COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH A/O TECHNOLOGY

		HYDRAULIC PLANT CAPACITY								
		0.91 ML/d			1.6 ML/d			3.2 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
COST ITEM	UNITS	1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	757 200	438 000	906 000	893 200	475 200	1 045 200	1 146 600	552 600	1 296 600
O&M	$/annum	27 100	20 100	32 300	31 700	20 700	39 100	30 300	13 300	41 200
	$/kg TP_R	20.39	15.13	20.69	13.57	8.90	14.24	6.69	2.86	7.50
Total Annual	$/annum	128 500	78 700	133 600	151 500	84 300	179 000	83 800	87 300	214 800
	$/kg TP_R	96.69	59.33	98.40	64.85	36.29	65.21	39.34	18.79	39.13

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

Retrofit of A/O technology to the model plants requires considerably less capital expenditure than either of the other totally biological nutrient control options (Bardenpho or UCT) in all cases. The A/O process configuration included in the evaluation was a two-stage system, requiring considerably less tankage, mixing, aeration and recycle pumping than the five-stage Bardenpho process or the three-stage UCT process. In this configuration, no nitrogen removal is achieved by the A/O process; in fact, nitrification in the aerobic section of the A/O system must be avoided to prevent inhibition of phosphorus release under anaerobic conditions. The A/O process retrofit also resulted in the lowest annual O&M costs of the three totally biological options, in all situations. The O&M costs associated with achieving nitrogen control in either Bardenpho or UCT systems is not a factor in the operation of the A/O process. In the conventional activated sludge plant at the largest hydraulic capacity, phosphorus removal to an effluent limit of 1.0 mg/L was achievable with an A/O retrofit including supplemental chemical addition at an incremental O&M cost of less than $1 per kg total phosphorus removed.

The total annual costs of retrofitting the model plants to the A/O process are compared to the costs of retrofitting chemical precipitation technology in Figure 22 (effluent total phosphorus limit 1.0 mg/L) and Figure 23 (effluent total phosphorus limit 0.3 mg/L). The cost advantage of supplemental chemical addition compared to effluent filtration is evident in Figure 22.

Based on total annual costs to achieve a 1.0 mg/L effluent total phosphorus limit, chemical phosphorus removal is more cost-effective than retrofitting to A/O technology in all situations except conventional activated sludge plants with hydraulic capacities of more than approximately 13.6 ML/d. Beyond this size, A/O becomes more cost-effective if supplemental chemical addition is practised in lieu of effluent filtration. Similarly, A/O becomes more cost-effective than chemical precipitation in conventional activated sludge plants with hydraulic capacities greater than approximately 13.6 ML/d when an effluent total phosphorus limit of 0.3 mg/L is imposed on the system. The annual O&M costs of all A/O retrofits to the model conventional activated sludge plants were lower than the comparable O&M costs of chemical precipitation technology irrespective of the method used to achieve an effluent total phosphorus limit of 1.0 mg/L. The A/O process was somewhat more capital intensive than the chemical precipitation technology in all situations, which affected the total annual costs.

A/O technology is patented by Air Products and Chemicals Inc. and licensing fees would be applicable to its use in Canada. These license fees, which are not included in the cost data presented in Figures 22 and 23, are a function of hydraulic capacity. At

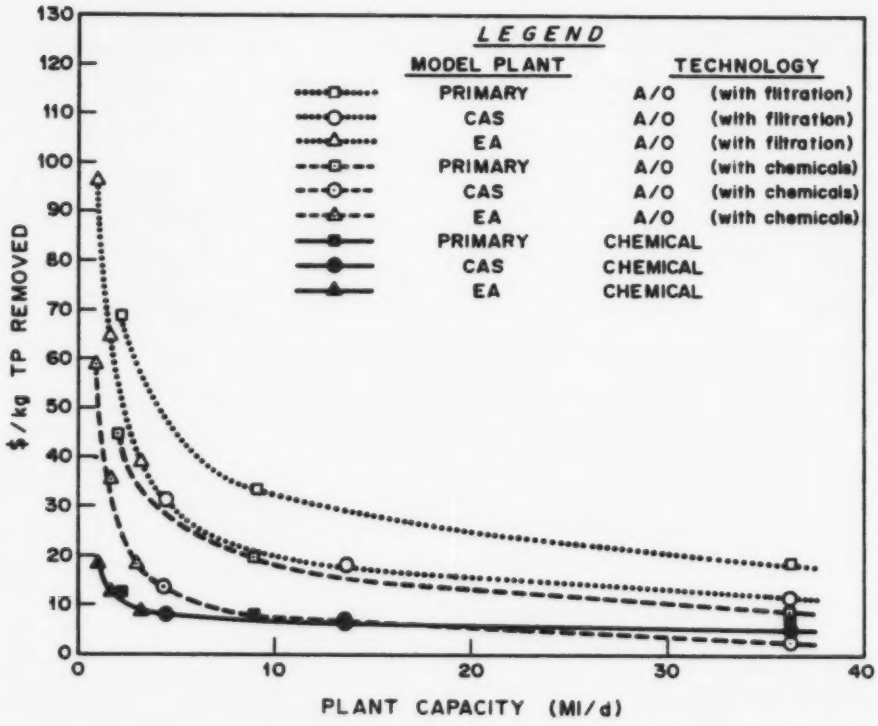


FIGURE 22 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR A/O AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS

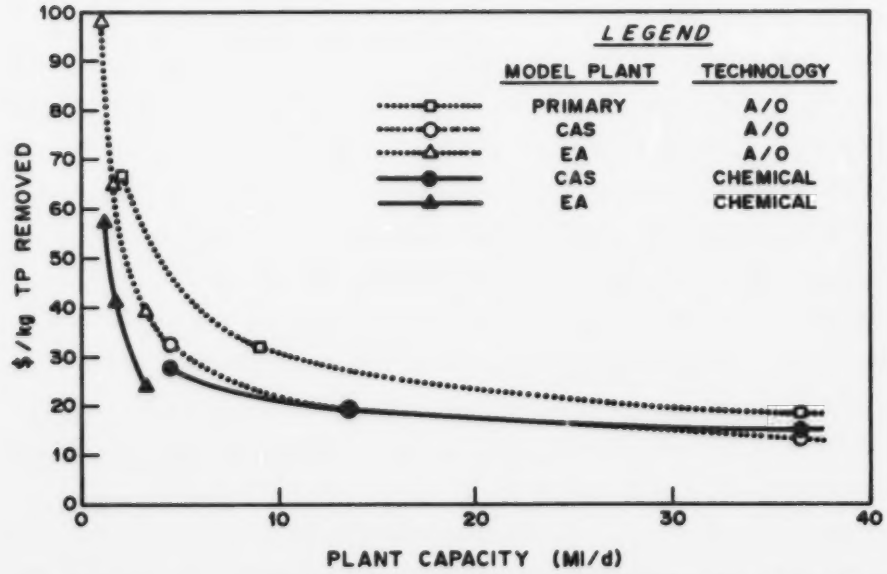


FIGURE 23 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR A/O AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS

an hydraulic flow of 36.4 ML/d, the license fee would increase the total annual cost of the A/O process retrofit by approximately $0.90 per kg phosphorus removed. Despite this additional cost, the A/O retrofit would maintain a slight cost advantage over chemical precipitation in this situation.

4.1.5 Phostrip Process. Cost summaries for model plant retrofits to the Phostrip process are presented in Tables 28, 29 and 30 for the model primary, conventional activated sludge and extended aeration plants, respectively. As noted in Section 3, in the case of Phostrip, neither effluent filtration nor supplemental chemical addition are required to achieve the 1.0 mg/L effluent total phosphorus limit. Effluent filtration is required to achieve the lower effluent total phosphorus limit of 0.3 mg/L.

Capital costs of retrofitting Phostrip technology to achieve phosphorus control are sensitive to the effluent total phosphorus limit imposed on the process. Capital requirements increased by approximately 50 to 100 percent as the effluent total phosphorus limit was reduced from 1.0 mg/L to 0.3 mg/L. The higher capital cost required to achieve the lower limit was primarily related to the addition of effluent filtration equipment to the Phostrip process flowsheet. In primary plant retrofit situations, the capital costs associated with Phostrip technology were higher than the costs associated with A/O technology but comparable to the costs of the Bardenpho process in most cases. Phostrip technology was less capital cost intensive than Bardenpho in secondary treatment plant retrofits (conventional activated sludge and extended aeration), but generally involved a higher capital cost than A/O process retrofits. In terms of annual O&M costs, the Phostrip retrofits were the least cost-effective of the EBPR technologies evaluated except in the small-capacity extended aeration plants.

The total annual costs of retrofitting the model plants to the Phostrip process are compared to the costs of retrofitting conventional chemical precipitation technology in Figure 24 (effluent total phosphorus limit 1.0 mg/L) and Figure 25 (effluent total phosphorus limit 0.3 mg/L). In all situations, the Phostrip retrofit was more expensive than the chemical precipitation retrofit to achieve the same effluent total phosphorus limit, on the basis of total annual costs. Similarly, annual O&M costs of the Phostrip retrofits were higher in all cases than the annual O&M costs associated with chemical precipitation, although the Phostrip O&M costs were less sensitive to the effluent total phosphorus limit than the O&M costs of conventional chemical precipitation technology.

Phostrip technology is licensed to Biospherics Inc.; a license fee would apply to its use in Canada. The license fee was not included in the costs calculated for the process retrofits.

TABLE 28 COSTS TO RETROFIT PRIMARY PLANTS WITH PHOSTRIP TECHNOLOGY

		HYDRAULIC PLANT CAPACITY					
		2.3 ML/d		9.1 ML/d		36.4 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	1 472 100	2 114 100	2 840 400	4 124 400	5 937 600	8 841 600
O&M	$/annum	124 000	140 500	280 700	311 700	871 000	943 000
	$/kg TP_R	37.42	36.08	21.15	19.99	16.41	15.12
Total Annual	$/annum	321 100	423 500	661 000	863 900	1 666 000	2 126 700
	$/kg TP_R	96.89	108.76	49.81	55.40	31.38	34.09

TABLE 29 COSTS TO RETROFIT CONVENTIONAL ACTIVATED SLUDGE PLANTS WITH PHOSTRIP TECHNOLOGY

		HYDRAULIC PLANT CAPACITY					
		4.5 ML/d		13.6 ML/d		36.4 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	867 000	1 750 200	1 377 600	2 997 600	2 101 200	5 005 200
O&M	$/annum	108 200	129 700	262 000	302 000	608 400	680 400
	$/kg TP_R	16.32	16.65	13.18	12.93	11.45	10.90
Total Annual	$/annum	224 300	364 000	446 400	703 300	889 700	1 350 500
	$/kg TP_R	33.84	46.74	22.45	30.10	16.74	21.63

TABLE 30 COSTS TO RETROFIT EXTENDED AERATION PLANTS WITH PHOSTRIP TECHNOLOGY

		HYDRAULIC PLANT CAPACITY					
		0.91 ML/d		1.6 ML/d		3.2 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	410 400	878 400	529 200	1 099 200	740 400	1 484 400
O&M	$/annum	39 000	46 000	49 500	60 500	82 000	99 000
	$/kg TP_R	29.35	29.47	21.19	22.04	17.55	18.03
Total Annual	$/annum	93 900	163 600	120 300	207 700	181 100	297 700
	$/kg TP_R	70.65	104.80	51.50	75.66	38.76	54.22

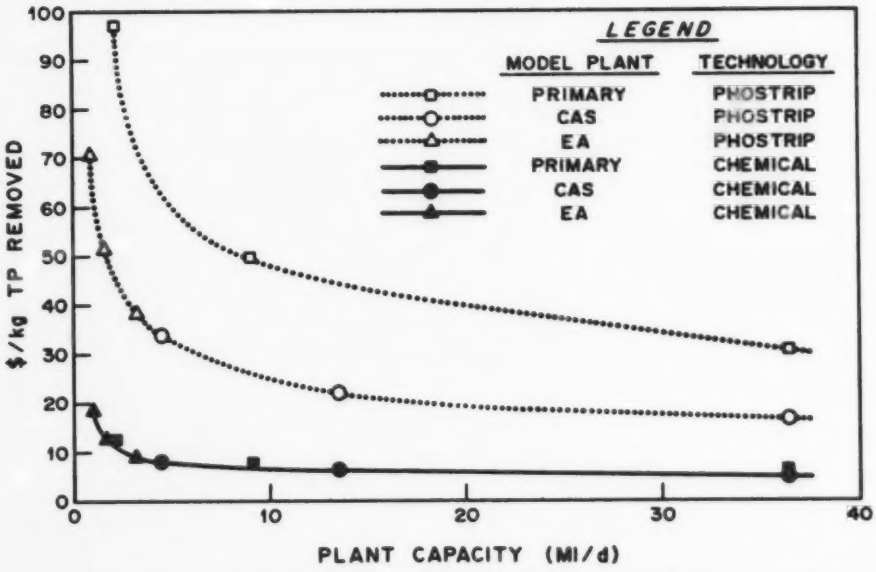


FIGURE 24 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR PHOSTRIP AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS

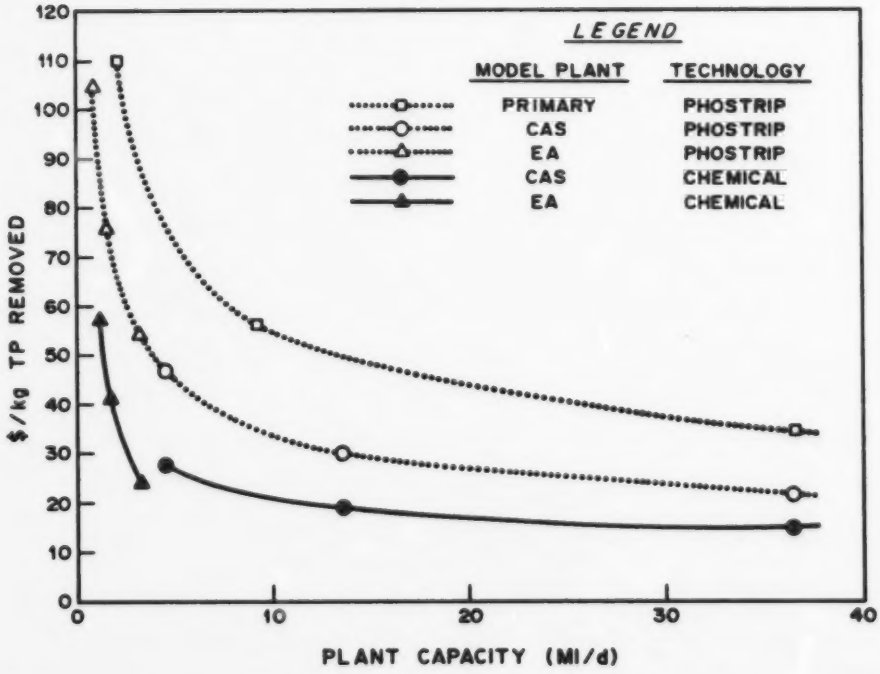


FIGURE 25 COMPARISON OF RETROFIT TOTAL ANNUAL COSTS FOR PHOSTRIP AND CHEMICAL PRECIPITATION TECHNOLOGIES TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS

4.1.6 Summary of Retrofit Costs. Total annual costs ($ per kg total phosphorus removed) for each retrofit option evaluated are presented in Tables 31, 32 and 33. Based on total annual costs, none of the EBPR processes are competitive with chemical precipitation in retrofit situations associated with an effluent total phosphorus limit of 1.0 mg/L, when effluent filtration is applied to Bardenpho, UCT and A/O retrofits. The most cost-effective EBPR process was more than twice as expensive as the conventional chemical approach. When supplemental chemical addition is used, the A/O process retrofit becomes the most cost-effective technology in conventional activated sludge plants at an hydraulic capacity of approximately 14 ML/d. All other EBPR options were at least twice as expensive, on a total annual cost basis, as chemical precipitation.

When the lower effluent total phosphorus limit (0.3 mg/L) is applied to the process retrofits, the EBPR options become more cost competitive, particularly at large scale in conventional activated sludge plant retrofit situations. However, only A/O is more cost-effective than chemical precipitation based on total annual costs.

Annual O&M costs ($ per kg total phosphorus removed) are presented in Tables 34, 35 and 36 for each retrofit option evaluated. As was the case in the total annual cost summaries, effluent filtration and supplemental chemical addition are dealt with separately. The O&M cost data show that all EBPR options with the exception of Phostrip represent lower annual O&M cost options than chemical precipitation in conventional activated sludge plants at an hydraulic capacity of 36.4 ML/d, even when effluent filtration is applied to achieve the 1.0 mg/L limit. These options are less costly on an annual O&M basis at a flow of 13.6 ML/d in conventional activated sludge plants when supplemental chemical addition is used to achieve 1.0 mg/L. At 36.4 ML/d, the O&M costs of A/O were less than 20 percent of the O&M costs of chemical precipitation.

In situations where an effluent total phosphorus limit of 0.3 mg/L is imposed, the annual O&M costs of all EBPR processes except Phostrip are less than the annual O&M costs of chemical precipitation in conventional activated sludge plants of 13.6 ML/d capacity. The A/O process was more cost-effective in this situation even at the lowest hydraulic flow capacity evaluated (4.5 ML/d) in conventional activated sludge plants.

4.2 New Plant Costs

Capital, annual O&M and total annual costs are calculated for new facilities achieving the selected effluent total phosphorus limits of 1.0 mg/L and 0.3 mg/L. The costs of the four EBPR process options (Bardenpho, UCT, A/O and Phostrip) were compared to the costs of a conventional activated sludge plant utilizing simultaneous

TABLE 31 SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)

		TOTAL ANNUAL COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			12.01			7.09		5.95
	Bardenpho*			95.61			46.14		27.59
	UCT*			97.16			45.22		26.07
	A/O*			68.86			33.32		18.37
	Phostrip			96.89			49.81		31.38
CAS	Chem					8.27		6.14	5.17
	Bardenpho*					46.83		28.12	18.46
	UCT*					49.83		28.41	18.33
	A/O*					31.31		18.51	11.81
	Phostrip					33.84		22.45	16.74
EA	Chem	18.28	12.41		8.20				
	Bardenpho*	117.38	84.04		51.77				
	UCT*	128.52	88.57		56.23				
	A/O*	96.69	64.85		39.34				
	Phostrip	70.65	51.50		38.76				

* Process incorporates final effluent filtration.

TABLE 32 SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)

		TOTAL ANNUAL COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			12.01			7.09		5.95
	Bardenpho*			70.71			32.44		19.39
	UCT*			72.27			31.52		17.87
	A/O*			44.00			19.62		10.17
	Phostrip			96.89			49.81		31.38
CAS	Chem					8.27		6.14	5.17
	Bardenpho*					28.84		16.28	10.28
	UCT*					31.85		16.58	10.15
	A/O*					13.34		6.68	3.62
	Phostrip					33.84		22.45	16.74
EA	Chem	18.28	12.41		8.20				
	Bardenpho*	80.07	55.57		31.29				
	UCT*	91.23	60.13		35.77				
	A/O*	59.33	36.29		18.79				
	Phostrip	70.65	51.50		38.76				

* Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 33 SUMMARY OF RETROFIT TOTAL ANNUAL COSTS TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT

		TOTAL ANNUAL COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			-			-		-
	Bardenpho			89.52			43.36		26.85
	UCT			90.86			42.77		25.56
	A/O			66.74			32.65		19.00
	Phostrip			108.76			55.40		34.09
CAS	Chem					28.03		19.22	14.46
	Bardenpho					45.60		27.80	19.08
	UCT					48.15		28.06	18.93
	A/O					32.38		19.63	13.39
	Phostrip					46.74		30.10	21.63
EA	Chem	57.14	40.44		28.44				
	Bardenpho	116.03	81.53		49.70				
	UCT	125.50	85.39		53.30				
	A/O	98.40	65.21		39.13				
	Phostrip	104.80	73.66		54.22				

TABLE 34 SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)

		TOTAL ANNUAL O&M COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			6.91			5.76		5.54
	Bardenpho*			30.63			14.62		8.16
	UCT*			33.80			15.52		8.25
	A/O*			22.78			10.85		6.37
	Phostrip			37.42			21.15		16.41
CAS	Chem					5.64		5.28	4.77
	Bardenpho*					10.20		6.26	3.93
	UCT*					11.95		6.97	4.04
	A/O*					5.60		3.34	2.17
	Phostrip					16.32		13.18	11.45
EA	Chem	5.79	5.26		4.62				
	Bardenpho*	24.90	18.71		10.34				
	UCT*	32.43	22.56		12.91				
	A/O*	20.39	13.57		6.49				
	Phostrip	29.35	21.19		17.55				

* Process incorporates final effluent filtration.

TABLE 35 SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)

		TOTAL ANNUAL O&M COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			6.91			5.76		5.54
	Bardenpho*			25.62			12.28		6.80
	UCT*			28.78			13.18		6.89
	A/O*			17.78			8.51		5.01
	Phostrip			37.42			21.15		16.41
CAS	Chem					5.64		5.28	4.77
	Bardenpho*					6.95		4.24	2.58
	UCT*					8.69		4.95	2.69
	A/O*					2.35		1.33	0.81
	Phostrip					16.32		13.18	11.45
EA	Chem	5.79	5.26		4.62				
	Bardenpho*	19.67	14.07		6.74				
	UCT*	27.20	17.94		9.32				
	A/O*	15.15	8.90		2.86				
	Phostrip	29.35	21.19		17.55				

* Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 36 SUMMARY OF RETROFIT ANNUAL O&M COSTS TO ACHIEVE 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT

		TOTAL ANNUAL O&M COSTS ($/kg TP removed)							
		HYDRAULIC PLANT CAPACITY (ML/d)							
MODEL PLANT	RETROFIT OPTION	0.91	1.6	2.3	3.2	4.5	9.1	13.6	36.4
PRIMARY	Chem			-			-		-
	Bardenpho			29.02			15.37		9.90
	UCT			31.72			16.14		9.98
	A/O			22.34			12.16		8.38
	Phostrip			36.08			19.99		15.12
CAS	Chem					10.63		9.18	7.90
	Bardenpho					11.75		8.27	6.30
	UCT					13.24		8.87	6.37
	A/O					7.83		5.79	4.77
	Phostrip					16.65		12.93	10.90
EA	Chem	13.07	11.18		6.28				
	Bardenpho	24.54	18.62		10.78				
	UCT	30.94	21.89		12.97				
	A/O	20.69	14.24		7.50				
	Phostrip	29.47	22.04		18.03				

chemical precipitation with ferric chloride. The cost analysis was done for plant sizes of 4.5 ML/d, 13.6 ML/d and 36.4 ML/d.

4.2.1 Chemical Precipitation. The capital, annual O&M and total annual costs for a new conventional activated sludge plant practising simultaneous chemical precipitation are presented in Table 37. Total annual costs associated with operation of the entire facility including the phosphorus removal component ranged from $47.68 to $107.35 per kg phosphorus removed depending on plant hydraulic capacity and the effluent phosphorus limit imposed on the process. Annual O&M costs represented approximately 40 to 50 percent of the total annual costs of the operation.

TABLE 37 NEW FACILITY COSTS - CONVENTIONAL ACTIVATED SLUDGE PLUS CHEMICAL ADDITION

		HYDRAULIC PLANT CAPACITY					
		4.5 ML/d		13.6 ML/d		36.4 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
COST ITEM	UNITS	1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	3 015 800	3 898 600	5 201 500	6 818 200	9 194 200	12 092 800
O&M	$/annum	269 800	315 200	607 600	718 300	1 300 686	1 540 600
	$/kg TP_R	40.65	40.42	30.52	30.71	24.50	24.70
Total Annual	$/annum	673 600	837 100	1 304 000	1 631 200	2 531 600	3 159 600
	$/kg TP_R	101.49	107.35	65.50	69.73	47.68	60.65

4.2.2 Bardenpho Process. The capital, annual O&M and total annual costs for a new Bardenpho facility of hydraulic capacity up to 36.4 ML/d are presented in Table 38. Costs associated with achieving the 1.0 mg/L effluent total phosphorus limit by effluent filtration and by supplemental chemical addition are included.

The capital costs associated with the new Bardenpho plant were approximately 20 to 50 percent higher than the capital costs of a conventional activated sludge plant achieving the same effluent phosphorus limit. Total annual costs associated with the Bardenpho facility ranged from $52.02 to $133.88 per kg phosphorus removed depending on plant capacity and the effluent phosphorus limit. In general, annual O&M costs represented a smaller percentage of the total annual costs (30 to 35 percent) for Bardenpho than for the conventional treatment facility.

4.2.3 UCT Process. Capital, annual O&M and total annual costs for new UCT facilities are presented in Table 39. The capital costs of the new UCT facility are approximately 5 to 10 percent lower than costs for a similar new Bardenpho facility. At plant capacities in excess of 13.6 ML/d, the annual O&M costs of the Bardenpho and UCT facilities were virtually identical, although the UCT facility was 5 to 10 percent less expensive to operate at the smallest scale. On a total annual cost basis, the UCT process was marginally less expensive than the Bardenpho process but more expensive at all sizes than the chemical precipitation alternative. Total annual costs for operation of new UCT facilities ranged from $50.83 to $128.15 per kg phosphorus removed.

4.2.4 A/O Process. Capital, annual O&M and total annual costs for new A/O facilities are presented in Table 40. Capital costs of new A/O facilities to achieve the lower effluent total phosphorus limit were lower than capital costs of new conventional plants at all sizes. In the case of A/O plants designed to achieve the higher effluent total phosphorus limit, those designed with effluent filters were more capital cost intensive than new conventional plants but those designed with supplemental chemical addition were less capital cost intensive than new conventional plants. Annual O&M costs of all new A/O facilities were lower, by as much as 10 percent, than annual O&M costs of new conventional plants with the single exception of the 4.5 ML/d plant designed to achieve a 1.0 mg/L effluent total phosphorus limit by filtration. On the basis of total annual costs, new A/O plants designed to achieve 0.3 mg/L total phosphorus were more cost-effective at all hydraulic capacities than the conventional plants and new A/O plants designed to achieve 1.0 mg/L total phosphorus by supplemental chemical addition were similarly more cost-effective than the comparable conventional plants. Total annual costs for operation of new A/O facilities ranged from $43.26 to $117.21 per kg total phosphorus removed. License fees for A/O technology, which are not included in Table 40, would add up to $0.90 per kg phosphorus removed to the total annual costs at a plant capacity of 36.4 ML/d, an increase in total annual cost of less than two percent.

4.2.5 Phostrip Process. New facility costs for Phostrip technology are summarized in Table 41. New Phostrip plants were considerably more capital intensive than new conventional plants at the same scale and achieving the same effluent phosphorus limit, exclusive of the additional licensing fees not included in the cost data. The capital costs of Phostrip technology were generally comparable to the capital costs of Bardenpho technology. Annual O&M costs of Phostrip plants were up to 15 percent higher than similar costs for new conventional plants. Total annual costs of new Phostrip plants

TABLE 38 NEW FACILITY COSTS - BARDENPHO TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY								
		4.5 ML/d			13.6 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
		1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	4 307 000	3 579 300	4 462 000	7 723 000	6 266 400	7 886 000	14 176 000	11 463 500	14 367 000
O&M	$/annum	312 000	290 500	336 000	651 000	611 000	720 000	1 229 000	1 227 000	1 483 000
	$/kg TP_R	47.01	43.77	43.07	32.71	30.69	30.78	24.44	23.11	23.78
Total Annual	$/annum	889 000	769 700	933 000	1 685 000	1 450 000	1 776 000	3 197 000	2 761 738	3 407 000
	$/kg TP_R	133.88	115.98	119.67	84.64	72.83	75.91	60.21	52.02	54.61

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 39 NEW FACILITY COSTS - UCT TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY								
		4.5 ML/d			13.6 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
		1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	4 187 000	3 459 300	4 342 000	7 477 000	6 020 400	7 640 000	13 660 000	10 947 500	13 851 000
O&M	$/annum	290 000	268 500	314 000	665 000	625 000	734 000	1 305 000	1 233 000	1 489 000
	$/kg TP_R	43.69	40.46	40.24	33.42	31.39	31.37	24.57	23.22	23.87
Total Annual	$/annum	850 000	731 600	895 000	1 666 000	1 431 000	1 757 000	3 133 000	2 698 700	3 344 000
	$/kg TP_R	128.15	110.24	114.79	83.69	71.88	75.10	59.02	50.83	53.60

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 40 NEW FACILITY COSTS - A/O TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY								
		4.5 ML/d			13.6 ML/d			36.4 ML/d		
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)								
		1.0*	1.0**	0.3	1.0*	1.0**	0.3	1.0*	1.0**	0.3
Capital	$	3 705 000	2 977 300	3 860 000	6 519 000	5 062 400	6 683 000	11 404 000	8 691 500	11 595 000
O&M	$/annum	282 000	260 500	306 000	593 000	553 000	662 000	1 205 000	1 133 000	1 389 000
	$/kg TP_R	42.49	39.25	39.22	29.80	27.78	28.30	22.69	21.34	22.27
Total Annual	$/annum	778 000	659 100	823 000	1 464 000	1 230 800	1 557 000	2 731 000	2 296 000	2 942 000
	$/kg TP_R	117.21	99.31	105.48	73.56	61.82	66.54	51.38	43.26	47.16

* Process incorporates final effluent filtration.
** Process incorporates supplemental chemical addition ($FeCl_3$).

TABLE 41 NEW FACILITY COSTS - PHOSTRIP TECHNOLOGY

COST ITEM	UNITS	HYDRAULIC PLANT CAPACITY					
		4.5 ML/d		13.6 ML/d		36.4 ML/d	
		EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)					
		1.0	0.3	1.0	0.3	1.0	0.3
Capital	$	4 000 000	4 900 000	6 500 000	8 150 000	10 431 000	13 335 000
O&M	$/annum	310 000	340 000	675 000	710 000	1 580 000	1 652 000
	$/kg TP_R	46.77	43.66	33.95	30.39	29.76	26.48
Total Annual	$/annum	840 000	960 000	1 600 000	1 820 000	2 977 000	3 438 000
	$/kg TP_R	126.74	123.27	80.46	77.89	56.07	55.10

ranged from $55.10 to $126.74 per kg total phosphorus removed depending on plant capacity and effluent phosphorus limit and exclusive of licensing fees.

4.2.6 Summary of New Plant Costs. Total annual costs for all process options designed to achieve an effluent total phosphorus limit of 1.0 mg/L are compared in Figures 26 and 27. Figure 26 compares costs for new EBPR plants incorporating final effluent filtration to achieve the required effluent quality, whereas Figure 27 shows costs for new EBPR plants incorporating supplemental chemical addition where required to achieve the effluent phosphorus limit. Figure 28 presents similar cost data for new facilities designed to achieve an effluent total phosphorus limit of 0.3 mg/L.

On the basis of total annual costs, the conventional plant (conventional activated sludge plus simultaneous chemical precipitation) is the most cost-effective option to achieve a 1.0 mg/L effluent total phosphorus limit over the range of hydraulic flows evaluated in the effluent filtration scenario. However, the A/O process becomes more cost-effective if supplemental chemical addition is provided as back-up to EBPR.

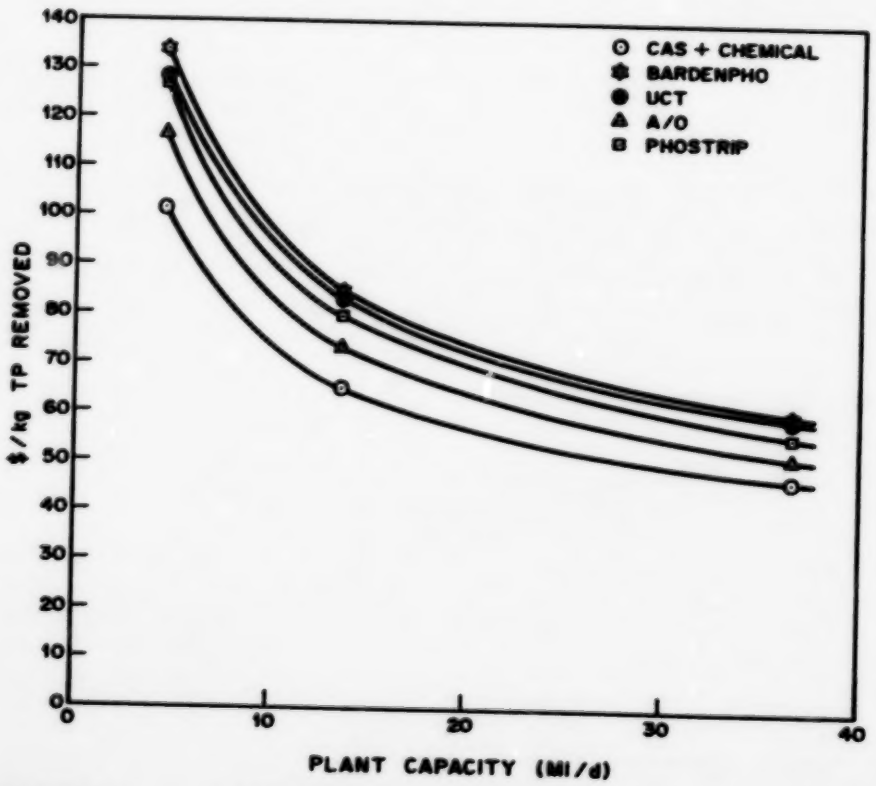


FIGURE 26 COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)

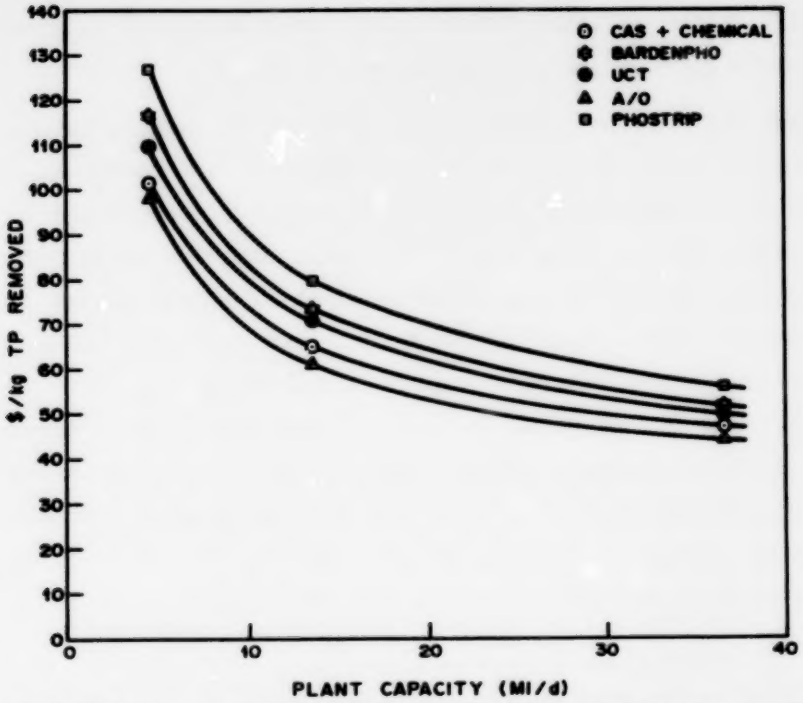


FIGURE 27 COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH CHEMICAL ADDITION)

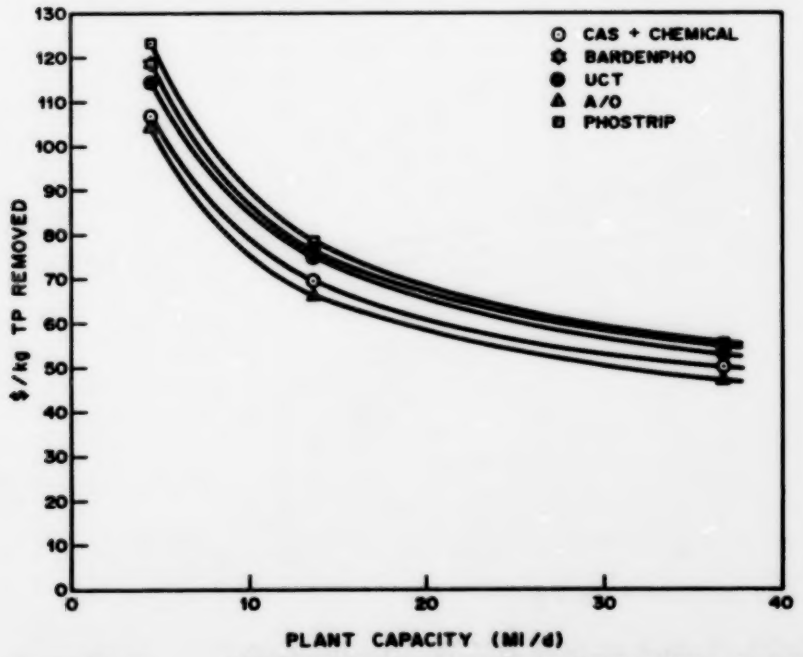


FIGURE 28 COMPARISON OF NEW PLANT TOTAL ANNUAL COSTS FOR 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT

Regardless of the approach taken to achieve the 1.0 mg/L effluent total phosphorus limit, both Bardenpho and UCT process are more expensive than the conventional technology based on total annual costs. However, both of these processes provide nitrogen control, which is not achieved by the conventional technology. When a more stringent effluent total phosphorus limit is imposed, A/O is more cost-effective than the chemical precipitation alternative. Under these circumstances, Phostrip is the most expensive process option based on total annual cost.

Annual O&M costs for the process options are presented in Table 42 (effluent limit 1.0 mg/L; filtration option), Table 43 (effluent limit 1.0 mg/L; supplemental chemical addition option) and Table 44 (effluent limit 0.3 mg/L). From these cost data, A/O is the lowest O&M cost option at all flows regardless of the effluent phosphorus limit with the single exception of the 4.5 ML/d plant designed to achieve 1.0 mg/L and incorporating effluent filtration. When the more stringent effluent total phosphorus limit is imposed, the annual O&M costs of all EBPR options except Phostrip are lower than the annual O&M costs of the conventional plant at flows in excess of approximately 13.6 ML/d.

4.3 Cost Sensitivity Analysis

The sensitivity of the costs of the conventional chemical precipitation technology to chemical dosage, chemical cost and chemical type (ferric chloride or alum) was evaluated. This sensitivity analysis was applied to the total annual costs and annual O&M costs of new conventional activated sludge plants practising simultaneous chemical precipitation to effluent total phosphorus limits of 1.0 mg/L and 0.3 mg/L.

TABLE 42 SUMMARY OF ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH FILTRATION)

	ANNUAL O&M COSTS ($/kg TP removed)		
	HYDRAULIC PLANT CAPACITY (ML/d)		
PROCESS	4.5	13.6	36.4
CAS & Chemical	40.65	30.52	24.50
Bardenpho*	47.01	32.71	24.44
UCT*	43.69	33.42	24.57
A/O*	42.49	29.80	22.69
Phostrip	46.77	33.95	29.76

* Process incorporates final effluent filtration.

TABLE 43 SUMMARY OF ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 1.0 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT (WITH SUPPLEMENTAL CHEMICAL ADDITION)

	ANNUAL O&M COST ($/kg TP removed)		
	HYDRAULIC PLANT CAPACITY (ML/d)		
PROCESS	4.5	13.6	36.4
CAS & Chemical	40.65	30.52	24.50
Bardenpho*	43.77	30.69	23.11
UCT*	40.46	31.39	23.22
A/O*	39.25	27.78	21.34
Phostrip	46.77	33.95	29.76

* Process incorporates supplemental chemical addition.

TABLE 44 SUMMARY OF ANNUAL O&M COSTS OF NEW PLANTS ACHIEVING 0.3 mg/L EFFLUENT TOTAL PHOSPHORUS LIMIT

	ANNUAL O&M COST ($/kg TP removed)		
	HYDRAULIC PLANT CAPACITY (ML/d)		
PROCESS	4.5	13.6	36.4
CAS & Chemical	40.42	30.71	24.70
Bardenpho	43.07	30.78	23.78
UCT	40.24	31.37	23.87
A/O	39.22	28.30	22.27
Phostrip	43.66	30.39	26.48

4.3.1 Effect of Chemical Dosage and Chemical Cost. Table 45 summarizes the effect of a 100 percent increase in chemical dosage on the total annual costs of a new conventional treatment facility. Also shown in Table 45 is the impact of a 100 percent increase in ferric chloride cost from $0.92 per kg Fe to $1.84 kg Fe on the total annual costs associated with the same facility.

Doubling the chemical dosage resulted in an increase in the total annual costs of between 4.2 percent and 12.0 percent, depending on plant hydraulic capacity and effluent total phosphorus limit. A similar increase in chemical cost produced an increase in the total annual costs of between 1.8 and 5.0 percent. The total annual costs were

TABLE 45 SENSITIVITY OF TOTAL ANNUAL COSTS TO CHEMICAL DOSAGE AND CHEMICAL COST

	PERCENT INCREASE IN TOTAL ANNUAL COST			
	HYDRAULIC PLANT CAPACITY			
	4.5 ML/d		36.4 ML/d	
	EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)			
FACTOR	1.0	0.3	1.0	0.3
100 Percent Increase in Dosage	4.2	8.3	6.0	12.0
100 Percent Increase in Cost	1.8	3.0	4.0	5.0

more sensitive to an increase in chemical dosage than an increase in chemical cost due to the higher sludge disposal costs associated with the increased sludge volumes generated at higher chemical dosages.

Data showing the impact of chemical dosage and chemical cost on plant annual O&M costs are presented in Table 46. As demonstrated for total annual costs, annual O&M costs are more sensitive to chemical dosage than to chemical cost. Annual O&M costs increased between 10.6 percent and 22.8 percent when chemical dosage was doubled compared to an increase of between 4.5 percent and 11.1 percent for a similar increase in chemical cost.

4.3.2 Effect of Chemical Type. The total annual costs associated with using alum in a new conventional activated sludge plant, at a cost of $220 per tonne (dry alum basis), were compared to the costs of using ferric chloride at a cost of $0.92 per kg Fe. The results of this analysis are summarized in Table 47.

4.4 Accuracy of Cost Estimates

The work of Pineau et al. (1985) demonstrated that, for six actual full-scale treatment facilities in Canada covering hydraulic capacities between 0.72 ML/d and 136 ML/d and a range of process configurations, CAPDET was capable of predicting construction costs to within approximately ±20 percent of actual costs. In the case of annual O&M costs, CAPDET was validated for 24 full-scale facilities, including primary plants, conventional activated sludge plants and aerated lagoons of capacities ranging

TABLE 46 SENSITIVITY OF ANNUAL O&M COSTS TO CHEMICAL DOSAGE AND CHEMICAL COST

	PERCENT INCREASE IN ANNUAL O&M COSTS			
	HYDRAULIC PLANT CAPACITY			
	4.5 (ML/d)		36.4 (ML/d)	
	EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)			
FACTOR	1.0	0.3	1.0	0.3
100 Percent Increase in Dosage	10.6	15.2	16.1	22.8
100 Percent Increase in Cost	4.5	6.8	7.5	11.1

TABLE 47 SENSITIVITY OF TOTAL ANNUAL COSTS TO CHEMICAL TYPE

	HYDRAULIC PLANT CAPACITY			
	4.5 (ML/d)		36.4 (ML/d)	
	EFFLUENT TOTAL PHOSPHORUS LIMIT (mg/L)			
FACTOR	1.0	0.3	1.0	0.3
Percent Increase in Total Annual Costs for Alum Use Compared to Ferric Chloride Use	3.0	2.5	6.0	5.0

from 1.2 ML/d to 313 ML/d, and found to be capable of predicting to within ±30 percent of actual O&M costs.

If these error envelopes are applied, for example, to the new plant costs for the various process options to achieve phosphorus removal, it would be concluded that the cost differences are, in general, within the error bounds of the estimating method and that no definitive decisions could be made with respect to the most cost-effective approach to phosphorus removal. However, it must be recognized that CAPDET was designed as a screening tool to rank process alternatives in terms of their cost-effectiveness. McGhee et al. (1983) concluded that as a screening tool, CAPDET was very

effective. The cost rankings of 20 alternatives based on CAPDET were essentially the same as the cost rankings developed by traditional manual costing methods, even though the specific costs developed by CAPDET did not necessarily agree with the costs developed by the manual methods. A major advantage of CAPDET is that costing techniques are applied consistently and it is assumed that cost estimating errors will generally be of the same type and in the same direction for each process alternative. Although the accuracy of the CAPDET capital and O&M cost predictions may be $\pm$20 percent and $\pm$30 percent of actual costs, the ranking and cost differences between alternatives should be predicted to a greater degree of accuracy.

Site-specific conditions at a particular plant will strongly influence the accuracy of the CAPDET predictions. As this cost assessment was conducted on a series of "model plants", these site-specific conditions would not affect the relative costs of the various alternatives.

5 POTENTIAL APPLICATIONS OF EBPR PROCESS RETROFITS IN CANADA

Factors other than economics affect the applicability of EBPR processes in retrofit situations. The limitations imposed on discharges of phosphorus and nitrogen, the wastewater characteristics, the types and sizes of existing facilities as well as the knowledge of and experience with EBPR processes compared to conventional chemical precipitation technology will also affect the final process selection. These and other factors are discussed in the following sub-sections with reference to the technical and economic factors already presented in Sections 2 and 4 of the report. As British Columbia, Alberta and Ontario are the only Canadian provinces where nutrient control requirements have been imposed to any major extent, and since these provinces account for the majority of the mechanical wastewater treatment facilities (primary and secondary) in Canada, the following discussions focus on the opportunities for EBPR process retrofits in these provinces.

5.1 Effluent Phosphorus Limits

Where phosphorus removal requirements have been imposed in Ontario and Alberta, the typical effluent concentration limit has been 1.0 mg/L, based on total phosphorus. In British Columbia, a limit of 1.5 mg/L total phosphorus has been the norm (Wetter, 1984). However, all provinces recognize that site-specific conditions may dictate more stringent requirements in terms of effluent total phosphorus limits. The International Joint Commission's Phosphorus Management Strategies Task Force has recommended that all future municipal point source discharges to the lower Great Lakes Basin consider total phosphorus removal requirements to concentrations in the order of 0.1 to 0.5 mg/L. Some Ontario municipal treatment facilities, including the City of Stratford WPCP, already have requirements to achieve more stringent effluent total phosphorus limits due to the sensitivity of the receiving waters.

The technical review of EBPR processes demonstrated that Bardenpho, UCT and A/O technologies were capable of achieving effluent total phosphorus concentrations of 1.0 to 2.0 mg/L without effluent polishing. Effluent total phosphorus limits of 1.0 mg/L or less would require final effluent filtration and/or supplemental chemical addition. In British Columbia, the EBPR processes alone would be technically feasible in most situations. In Ontario and Alberta, EBPR processes would require either effluent filtration or supplemental chemical addition to consistently achieve the 1.0 mg/L effluent limit. Effluent filtration was shown to be capital cost intensive, eliminating any O&M

cost advantage associated with the EBPR processes. Where an effluent total phosphorus limit of 1.0 mg/L is imposed, Bardenpho, UCT and A/O are only cost competitive with conventional chemical precipitation technology if supplemental chemical addition is practised. Phostrip technology was not cost competitive with conventional technology achieving a 1.0 mg/L effluent total phosphorus limit even though effluent filtration is not required.

The economic assessment demonstrated that, when a more stringent effluent total phosphorus limit is imposed, such as has been recommended for numerous plants in Ontario by the IJC, the EBPR technologies become more cost competitive with the chemical precipitation process. In fact, in conventional activated sludge plants with hydraulic capacities of 36.4 ML/d, A/O was the least cost retrofit option. The imposition of effluent total phosphorus limits of less than 1.0 mg/L will favour the implementation of EBPR processes because, under these circumstances, EBPR technologies have an O&M cost advantage over the conventional chemical precipitation technology, particularly in conventional activated sludge plants.

5.2 Effluent Nitrogen Limits

Nitrogen control requirements have been imposed in Canada only on a site-specific basis in situations where the assimilative capacity of the receiving waters is limited. Typically, nitrogen limitations have been based on effluent ammonia concentrations to control dissolved oxygen depletion in the receiving waters due to the nitrogenous demand and to prevent ammonia toxicity to the receiving water biota. Requirements to achieve complete nitrogen control, which implies the implementation of denitrification, have not been widespread. To our knowledge, the only municipal treatment facilities presently required to achieve a total nitrogen limit are Orangeville, Ontario, and Kelowna, British Columbia.

EBPR processes such as Bardenpho and UCT provide complete nitrogen control and, in fact, denitrification is a process requirement to ensure optimum phosphorus removal biologically. The higher capital and O&M costs of Bardenpho and UCT systems are associated with the additional costs of nitrifying and denitrifying the wastewater. Additional reactor volume and oxygen transfer capacity must be provided. In return for these higher costs, Bardenpho and UCT produce a higher quality effluent in terms of the total nutrient load (nitrogen and phosphorus) than the other process options. Where nitrogen control is not a requirement, this benefit in effluent quality may not justify the higher cost of the technology.

The imposition of a nitrogen control requirement on an existing treatment facility will significantly affect the cost analysis presented. Under these circumstances, the costs of the conventional technology will increase and may exceed the costs of the Bardenpho and UCT options. This cost analysis was not within the scope of this study. However, there is no doubt that in situations where nitrogen control requirements are imposed, the EBPR options that incorporate nitrogen control (Bardenpho, UCT and A^2/O) will become more favourable from a cost standpoint.

The cost analyses presented were based on a five-stage Bardenpho process configuration and a two-stage A/O process configuration. Bardenpho systems can be designed in a three-stage configuration to meet less stringent effluent nitrogen limits. A/O systems can be designed in a three-stage configuration in situations where denitrification is required. The process configuration will have a significant impact on the relative capital and O&M costs of the individual EBPR options.

5.3 Wastewater Characteristics

EBPR processes, particularly those that incorporate nitrogen control, are more sensitive to the characteristics of the raw wastewater than conventional chemical precipitation processes. As discussed in Section 2, the EBPR processes require that the COD (BOD_5), TKN and phosphorus concentrations be present at certain minimum ratios (COD/TKN $\geq$ 12 for Bardenpho, UCT and A^2/O; BOD_5/soluble P > 10 for A/O). The wastewater characteristics established for this investigation (Section 1.2) would be marginal for design of an EBPR process due to the low COD/TKN ratio of 10. Supplementation of the influent COD with volatile fatty acids from an acid digester or sludge thickener, along with at least partial bypassing of the primary clarifier, might be required.

Ontario has the most complete data base for wastewater characteristics available in Canada. Based on 1980 data for 90 Ontario water pollution control plants, the wastewater characteristics presented in Table 48 were established. Assuming a BOD_5/COD ratio of 0.4, the average COD/TKN ratio in Ontario wastewater is 9.3 and 50 percent of Ontario plants have a COD/TKN ratio of less than 7.8. Primary clarification would further reduce the COD/TKN ratio in the influent to the biological section of the plant as COD is removed to a greater extent in the primaries than TKN.

Further, wastewater in Ontario has a relatively low total phosphorus concentration as a result of the regulation of the phosphorus content of detergents. EBPR technology was largely developed in South Africa, where raw wastewater strengths are

TABLE 48 CHARACTERISTICS OF RAW WASTEWATERS IN ONTARIO

Parameter	Concentration (mg/L)	
	Mean	Median
Total BOD_5	120.	95.
TSS	180.	140.
Total P	6.8	5.8
TKN	32.2	30.6
NH_3-N	20.1	18.2

considerably higher than in Canada. Under these circumstances, the cost advantages of EBPR processes are more readily demonstrated. The O&M costs associated with conventional chemical precipitation technology are related to the chemical dosage requirement, which is a function of the influent phosphorus level. The cost sensitivity analysis demonstrated that a 100 percent increase in chemical dosage would result in an increase in annual O&M costs of as much as 22.8 percent in new plants, depending on plant size and effluent phosphorus limit. On the other hand, the O&M costs associated with totally biological EBPR processes are relatively independent of influent phosphorus concentration.

Wastewater characteristics in Canada, in general, do not favour the implementation of EBPR processes. More than 50 percent of Ontario plants would not be suitable for EBPR without supplementation of the influent COD, which would increase the cost of the EBPR process and add further complexity to the operation. In addition, the low raw wastewater phosphorus concentrations common in Canada reduce the cost advantages associated with EBPR processes.

5.4 Characteristics of Existing Canadian WPCPs

The economic analysis demonstrated that EBPR processes are most cost-effective in retrofit situations where the existing facility is a conventional activated sludge plant with a hydraulic flow in excess of approximately 13.6 ML/d. EBPR processes cannot compete economically with chemical precipitation in primary plants required to achieve an effluent limit of 1.0 mg/L. The majority of extended aeration plants are too small to provide an economic advantage to EBPR processes and, as packaged plants, are not particularly amenable to an EBPR process retrofit.

Ontario has the most complete data base on plant design and capacity in Canada. In Ontario, there are 44 conventional activated sludge plants with design

hydraulic capacities of 13.6 ML/d or greater; however, the actual hydraulic loading on conventional activated sludge plants in Ontario is only 69 percent of design, on average. Therefore, some of the 44 plants may be operating at flows less than 13.6 ML/d.

In Alberta, there are only five conventional activated sludge plants, ranging in size from 18 ML/d to 450ML/d. Only two plants (Fish Creek and Bonnybrook in Calgary) are required to achieve phosphorus removal. Influent phosphorus concentrations are approximately 4.5 mg/L and the effluent requirement (1.0 mg/L) is consistently achieved by alum addition at a dosage of approximately 3 mg/L (Spink, 1984).

British Columbia has Canada's only operating full-scale EBPR process, at Kelowna in the Okanagan Valley. Other major plants currently practicing phosphorus removal include Penticton (18 ML/d), Kamloops (36.4 ML/d), Salmon Arm (2.7 ML/d) and Whistler (expanding to 9 ML/d). All but the Kamloops plant are conventional activated sludge plants, which achieve phosphorus removal by chemical precipitation using alum or pickle liquor. Pickle liquor is obtained as a waste product from Vancouver free of charge except for transportation costs (Wetter, 1984). Based on 1977 data, there were 21 activated sludge plants in British Columbia, of which 7 were high-rate activated sludge plants. Of the 14 conventional plants, only 3 would appear to have a design hydraulic capacity in excess of 13.6 ML/d.

EBPR processes are more susceptible to hydraulic and strength variations in the raw wastewater than chemical precipitation processes. Flow equalization has been provided in many of the existing EBPR plants and appears to improve the consistency of process operation. Few existing Canadian WPCPs have flow equalization capabilities. The provision of such facilities in an EBPR process retrofit would increase the capital costs beyond those presented in Section 4. Newer examples of EBPR technology have also demonstrated the importance of process control and the need to minimize re-aeration of the wastewater upstream of the anaerobic reactor. In new plants, incorporation of these details in the design will not significantly affect the overall plant costs. However, in retrofit situations, implementation of dissolved oxygen control and modifications to plant hydraulics to minimize re-aeration can represent significant cost items depending on plant hardware and configuration. The site-specific factors will have an impact on the cost-effectiveness of EBPR processes in conventional activated sludge plants that appear to be suitable for EBPR process retrofit on the basis of size and wastewater characteristics. It is important to note that this cost analysis was done on model plants where site-specific factors were not considered.

5.5 Sludge Management Implications

EBPR processes are sensitive to sludge handling processes applied to phosphorus-rich waste biological solids. Anaerobic conditions favour the release of phosphorus from the sludge. Recycle of anaerobic digester supernatant to the treatment plant should be avoided and waste activated sludge thickening should be done by dissolved air flotation.

At the Bardenpho facility in Kelowna, waste activated sludge is thickened by dissolved air flotation and combined with gravity-thickened primary sludge for off-site composting. In Ontario, most conventional activated sludge plants practise two-stage anaerobic digestion and digester supernatant recycle, with disposal of sludge to agricultural land. Province of Ontario guidelines for sewage sludge utilization on agricultural lands state that only 'processed' sludges are acceptable, where 'processed' sludge is defined as a sludge that has been treated by aerobic or anaerobic digestion, or by other means of stabilization. Few Ontario plants practise waste sludge thickening.

The implementation of EBPR processes may require major changes to sludge management practices in an existing plant. The best approach to sludge handling will depend on the existing hardware, the existing disposal methodology and the local regulatory requirements. In this cost analysis, the EBPR processes were not unduly handicapped by the specification of capital and O&M intensive sludge handling processes. However, in any specific retrofit situation, sludge management may significantly affect the costs of EBPR processes.

5.6 State of Development of Phosphorus Removal Technologies

Phosphorus removal by simultaneous chemical precipitation has been widely applied in Canada for more than ten years. It has been shown to be a practical, simple technology that will consistently provide compliance with effluent requirements. Understanding of the process chemistry has developed to the point that the technology can be implemented without extensive laboratory, pilot-scale or full-scale testing. Chemical dosage relationships are well defined and the impact on sludge generation rates can be readily predicted from past experience.

EBPR processes are relatively new to North America and, as demonstrated by the performance data presented in Section 2, few examples of the technology have been capable of consistently meeting their design objectives. Design relationships are generally empirical and the biochemistry of the phosphorus-removal mechanisms are not completely understood. Provisions for supplemental chemical addition and/or effluent

filtration are generally necessary unless extensive treatability data from pilot or full-scale tests are available to demonstrate process reliability. Even with extensive treatability data, supplemental chemical feed equipment would be recommended as backup to the EBPR process. EBPR processes are considerably more complex than the conventional chemical precipitation technology, requiring increased operator knowledge, training and supervision and/or the implementation of on-line monitoring and control equipment. These factors may outweigh the economic advantages that have been shown to apply to EBPR processes in some situations.

5.7 Avenues of Capital Funding

The economic analysis demonstrated that, in general, EBPR processes are more capital intensive than the conventional chemical precipitation technology, but result in lower annual O&M costs in conventional activated sludge retrofit situations. Overall, the total annual costs of the EBPR processes are higher because the additional costs of financing the capital expenditure outweigh the advantages in terms of O&M costs.

This cost analysis was based on amortization of the capital costs at an annual interest rate of 12 percent over a term of 20 years. Because of the higher capital costs associated with the EBPR processes retrofits, the economics of the EBPR options are more sensitive to the assumptions made with respect to capital funding than the conventional technologies. In many situations, grants or low-interest loans are provided to a municipality to cover part or all of the capital costs of treatment works. The availability of capital funding assistance will have a significant impact on the cost-effectiveness of EBPR processes.

5.8 Summary

Given the present situation in Canada in terms of nutrient limits and plant size, the number of existing treatment facilities that would be candidates for a retrofit of an EBPR technology is probably less than 15. Circumstances that would lead to more widespread implementation of these technologies include:

i) the imposition of effluent total phosphorus limits of 0.5 mg/L or less on a widespread basis;

ii) the imposition of nitrogen removal requirements on existing municipal treatment facilities;

iii) major increases in the cost of ferric chloride in Ontario;

iv) the development of greater confidence among users and regulatory agencies in the

capability of full-scale EBPR processes in North America to consistently achieve effluent total phosphorus limits of 1.0 mg/L or less; and

v) capital funding assistance.

It also appears, based on the information compiled during the course of this investigation, that EBPR processes and chemical precipitation processes should be considered complementary rather than as completely separate process options. EBPR facilities should be designed to biologically reduce the phosphorus content of the wastewater to a practical minimum and the residual phosphorus should be removed chemically to the prescribed effluent target. The chemical feed system should be designed to handle the total influent phosphorus load on the plant to ensure that effluent requirements are met consistently.

6 CONCLUSIONS

The technical review and economic evaluations conducted during the course of this investigation lead to the following conclusions with respect to biological phosphorus removal technologies and their potential application in Canada.

6.1 Technical Feasibility of EBPR Processes

1) Of the EBPR processes evaluated in detail in this report, three (Bardenpho, A/O and Phostrip) are reasonably well developed and full-scale examples of each are operating in North America. Newer facilities based on these technologies have showed more consistent performance as a result of improved engineering design and a better understanding of the mechanisms involved in EBPR processes. The UCT process remains to be demonstrated at full-scale.

2) All four EBPR processes are capable of producing a final effluent containing between 1.0 and 2.0 mg/L total phosphorus. In situations where the effluent total phosphorus limit is 1.0 mg/L, Bardenpho, UCT and A/O require either effluent filtration or supplemental chemical addition to ensure compliance. Both effluent filtration and supplemental chemical addition must be provided in conjunction with Bardenpho, UCT and A/O processes to achieve an effluent total phosphorus limit of 0.3 mg/L. The Phostrip process can achieve an effluent total phosphorus limit of 1.0 mg/L without effluent filtration, but filtration is required to achieve an effluent total phosphorus limit of 0.3 mg/L consistently.

3) EBPR and chemical precipitation should be considered as complementary processes. Ideally, EBPR facilities should be designed to biologically reduce the phosphorus content of the wastewater to a practical minimum and the residual phosphorus should be removed chemically to the prescribed effluent limit. Chemical back-up to the EBPR system is recommended.

4) Sludge management practices significantly affect the performance of EBPR processes and information with respect to sludge management at EBPR facilities is sparse. More attention needs to be devoted to this area.

6.2 Economics of EBPR Process Retrofits

1) Based on total annual costs, none of the EBPR processes were cost competitive with chemical precipitation to an effluent total phosphorus limit of 1.0 mg/L when effluent filtration was practised. When supplemental chemical addition was

practised, the A/O process became the most cost-effective technology in conventional activated sludge plants at hydraulic capacities above approximately 14 ML/d. All other EBPR options were at least twice as expensive as chemical precipitation in retrofit situations when the effluent limit was 1.0 mg/L total phosphorus.

2) In situations where the effluent total phosphorus limit was 0.3 mg/L, the EBPR options were more cost competitive with chemical precipitation, particularly in larger conventional activated sludge plant retrofits. However, only A/O was more cost-effective than chemical precipitation based on total annual costs.

3) The annual O&M costs of all EBPR options with the exception of Phostrip were lower than O&M costs of chemical precipitation in conventional activated sludge plant retrofits at an hydraulic capacity of 36.4 ML/d, regardless of effluent total phosphorus limit.

6.3 Economics of New EBPR Facilities

1) On the basis of total annual costs, the conventional simultaneous chemical precipitation technology was the most cost-effective option to achieve a 1.0 mg/L effluent total phosphorus limit if effluent filtration was provided. A/O would be more cost-effective if supplemental chemical addition were provided or if a lower effluent phosphorus limit were imposed.

2) In terms of annual O&M costs, A/O was the lowest cost option at all flows regardless of the effluent phosphorus limit with the single exception of the 4.5 ML/d plant designed to achieve 1.0 mg/L and incorporating effluent filtration.

3) When a 0.3 mg/L effluent total phosphorus limit was imposed, all EBPR processes except Phostrip had lower annual O&M costs than the conventional plant when hydraulic capacity exceeded 13.6 ML/d.

6.4 Applicability of EBPR Process Retrofits in Canada

1) Given the present situation in Canada in terms of effluent requirements, existing facilities and wastewater characteristics, the number of wastewater treatment plants that would be candidates for a retrofit of EBPR technology is very limited.

2) Circumstances that would lead to more widespread implementation of EBPR technologies include:

   i) the widespread imposition of effluent total phosphorus limits of 0.5 mg/L or less;

ii) the imposition of nitrogen removal requirements;

iii) major increases in chemical costs; and

iv) the development of greater confidence in the performance of EBPR processes.

3) Site-specific factors can have a significant impact on the relative costs of EBPR and chemical precipitation processes in actual plants. The sensitivity of the retrofit costs to these specific factors needed to be evaluated in further detail.

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APPENDIX

COST FUNCTIONS FOR PROCESS RETROFITS

INDEX OF RETROFIT EQUIPMENT COST CURVES

Figure Number	Cost Function	Retrofit Application		
		Model Plant Type *	P Removal Process	TP Effluent Limit (mg/L)
A-1	Pumping Station - Capital	Primary	Bardenpho UCT A/O	0.3, 1.0 0.3, 1.0 0.3, 1.0
A-2	Pumping Station - O&M	Primary	Bardenpho UCT A/O	0.3, 1.0 0.3, 1.0 0.3, 1.0
A-3	Basin Transfer Piping - Capital	Primary & CAS	Bardenpho UCT A/O	0.3, 1.0 0.3, 1.0 0.3, 1.0
A-4	New A/O Basin - Capital	Primary	A/O	0.3, 1.0
A-5	CAS Basin Retrofit for A/O - Capital	CAS	A/O	0.3, 1.0
A-6	EA Basin Retrofit for A/O - Capital	EA	A/O	0.3, 1.0
A-7	A/O Process Mixers - Capital	All	A/O	0.3, 1.0
A-8	A/O Process Blower Facilities - Capital	All	A/O	0.3, 1.0
A-9	A/O Basin - O&M	All	A/O	0.3, 1.0
A-10	Final Effluent Filter - Capital	All	Bardenpho UCT A/O Phostrip Chemical	0.3, 1.0 0.3, 1.0 0.3, 1.0 0.3 0.3
A-11	Final Effluent Filter - O&M	All	Bardenpho UCT A/O Phostrip Chemical	0.3, 1.0 0.3, 1.0 0.3, 1.0 0.3 0.3

* CAS = Conventional Activated Sludge; EA = Extended Aeration

INDEX OF RETROFIT EQUIPMENT COST CURVES (Cont'd)

Figure Number	Cost Function	Retrofit Application		
		Model Plant Type *	P Removal Process	TP Effluent Limit (mg/L)
A-12	Chemical P Removal - Capital	All	Bardenpho UCT A/O Chemical	0.3, 1.0 0.3, 1.0 0.3, 1.0 0.3, 1.0
A-13	Chemical P Removal - O&M	All	Chemical	0.3, 1.0
A-14	New Bardenpho Basin - Capital	Primary	Bardenpho	0.3, 1.0
A-15	CAS Basin Retrofit for Bardenpho - Capital	CAS	Bardenpho	0.3, 1.0
A-16	EA Basin Retrofit for Bardenpho - Capital	EA	Bardenpho	0.3, 1.0
A-17	Bardenpho Process Mixers - Capital	All	Bardenpho	0.3, 1.0
A-18	Bardenpho/UCT Blower Facilities - Capital	All	Bardenpho UCT	0.3, 1.0 0.3, 1.0
A-19	Bardenpho/UCT Basin - O&M	All	Bardenpho UCT	0.3, 1.0 0.3, 1.0
A-20	Bardenpho Internal Recycle - Capital	All	Bardenpho	0.3, 1.0
A-21	Bardenpho Internal Recycle - O&M	All	Bardenpho	0.3, 1.0
A-22	CAS Aeration Basin - Capital	Primary	Phostrip	0.3, 1.0
A-23	CAS Aeration Basin - O&M	Primary	Phostrip	0.3, 1.0
A-24	Secondary Clarifier - Capital	Primary	Phostrip	0,3 1.0
A-25	Secondary Clarifier - O&M	Primary	Phostrip	0.3, 1.0
A-26	Phostrip Stripper tank - Capital	All	Phostrip	0.3, 1.0
A-27	Phostrip Stripper Tank - O&M	All	Phostrip	0.3, 1.0

* CAS = Conventional Activated Sludge; EA = Extended Aeration

INDEX OF RETROFIT EQUIPMENT COST CURVES (Cont'd)

Figure Number	Cost Function	Retrofit Application		
		Model Plant Type *	P Removal Process	TP Effluent Limit (mg/L)
A-28	Phostrip Underflow Piping - Capital	All	Phostrip	0.3, 1.0
A-29	Phostrip Underflow Piping - O&M	All	Phostrip	0.3, 1.0
A-30	Phostrip Reactor-Clarifier - Capital	All	Phostrip	0.3, 1.0
A-31	Phostrip Reactor-Clarifier - O&M	All	Phostrip	0.3, 1.0
A-32	Lime Feed System - Capital	All	Phostrip	0.3, 1.0
A-33	Lime Feed System - O&M	All	Phostrip	0.3, 1.0
A-34	New UCT basin - Capital	Primary	UCT	0.3, 1.0
A-35	CAS Basin Retrofit for UCT - Capital	CAS	UCT	0.3, 1.0
A-36	EA Basin Retrofit for UCT - Capital	EA	UCT	0.3, 1.0
A-37	UCT Process Mixers - Capital	All	UCT	0.3, 1.0
A-38	UCT Internal Recycle - Capital	All	UCT	0.3, 1.0
A-39	UCT Internal Recycle - O&M	All	UCT	0.3, 1.0
B-1	Flow Equalization - Capital	Supplemental Cost Curve		
B-2	Flow Equalization - O&M	Supplemental Cost Curve		
B-3	Chlorine Contact Chamber - Capital	Supplemental Cost Curve		
B-4	Chlorine Contact Chamber - O&M	Supplemental Cost Curve		
B-5	Dissolved Air Flotation Thickener - Capital	Supplemental Cost Curve		
B-6	Dissolved Air Flotation Thickener - O&M	Supplemental Cost Curve		

* CAS = Conventional Activated Sludge; EA = Extended Aeration

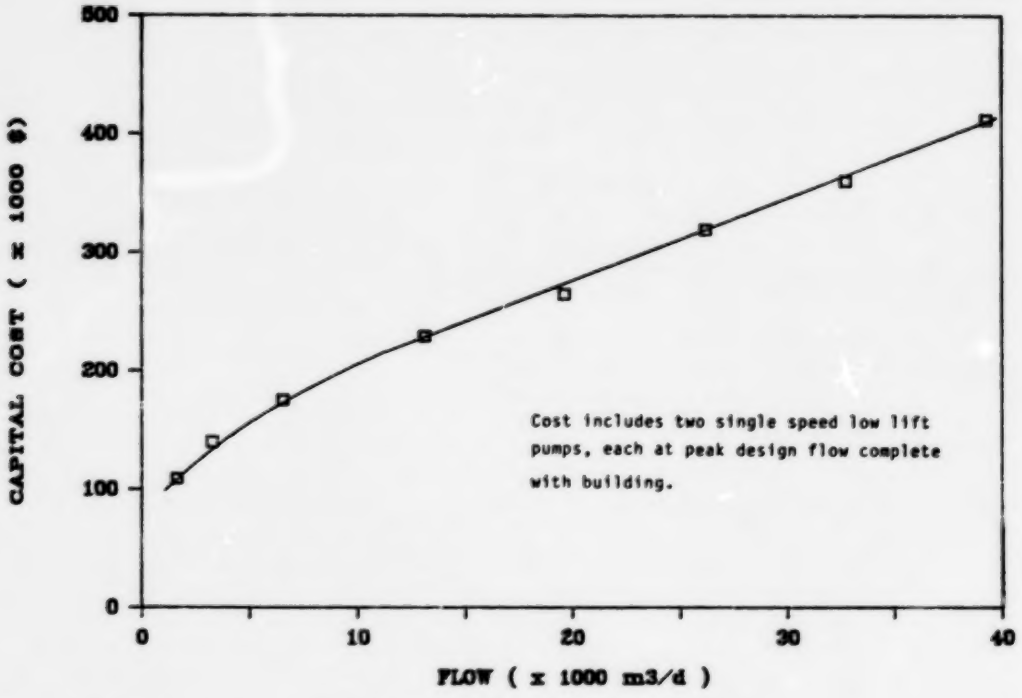

FIGURE A-1 PUMPING STATION COST - CAPITAL

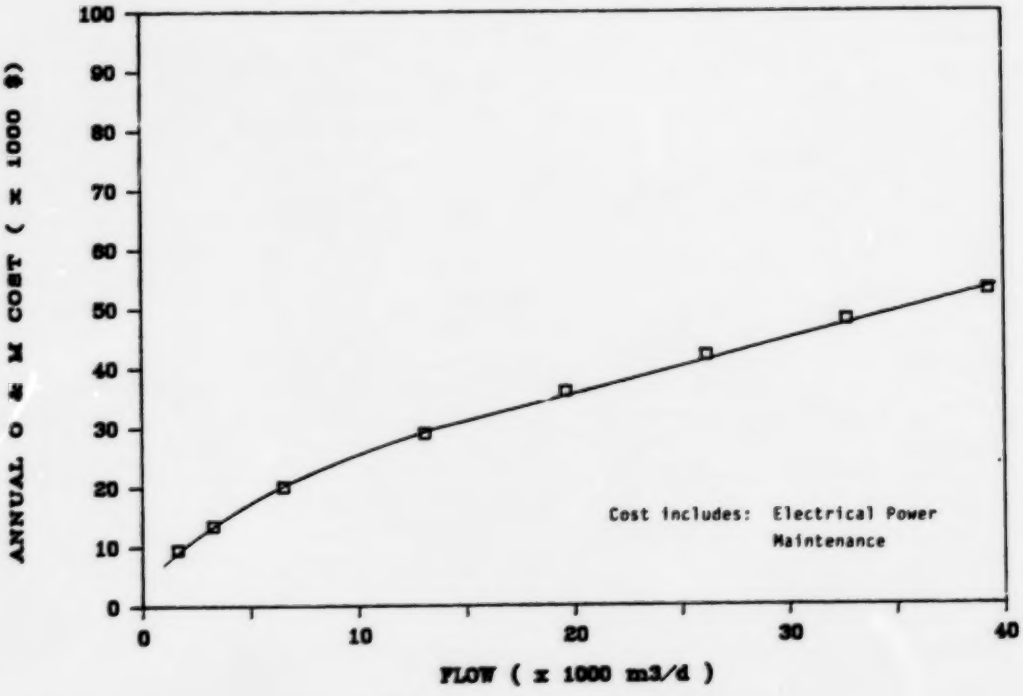

FIGURE A-2 PUMPING STATION COST - O&M

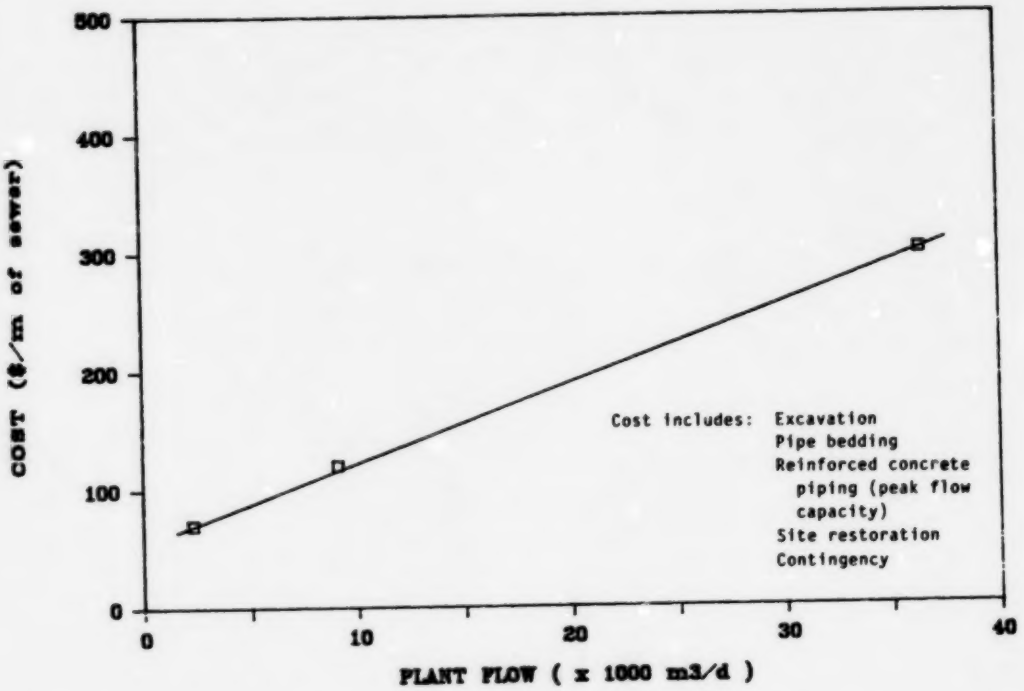

FIGURE A-3 BASIN TRANSFER PIPING COST - CAPITAL

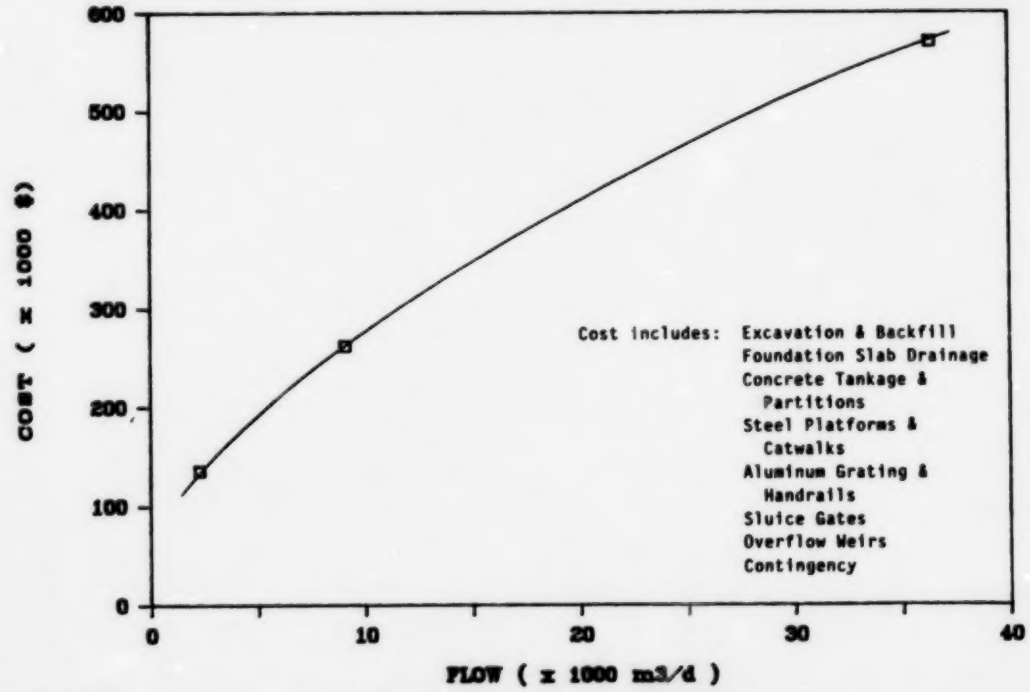

FIGURE A-4 NEW A/O BASIN COST - CAPITAL

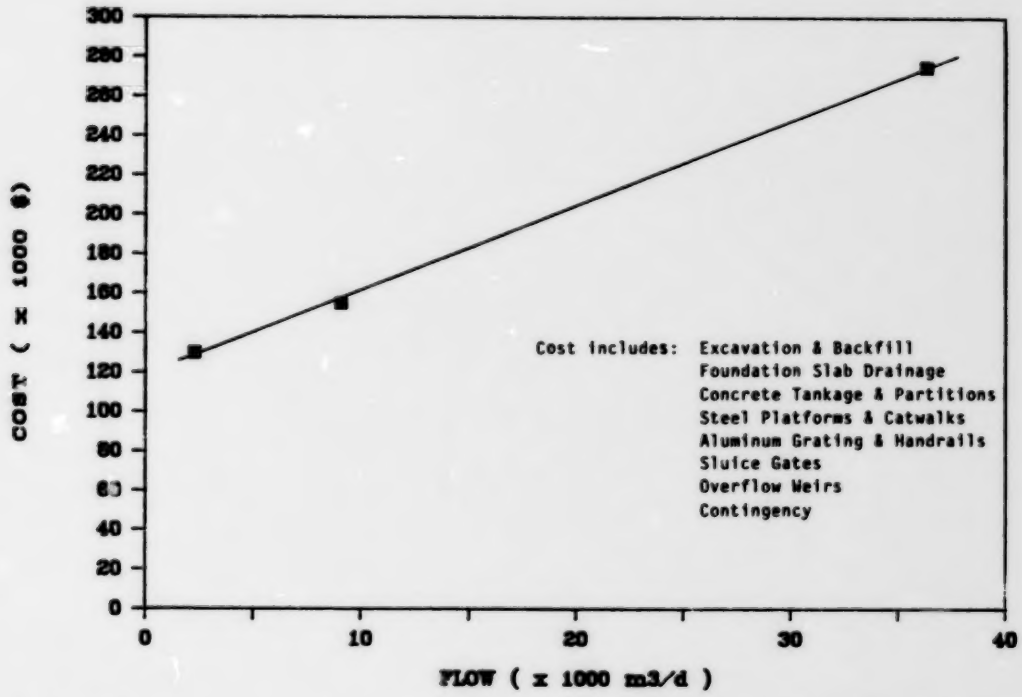

FIGURE A-5 CAS AERATION BASIN RETROFIT TO A/O COST - CAPITAL

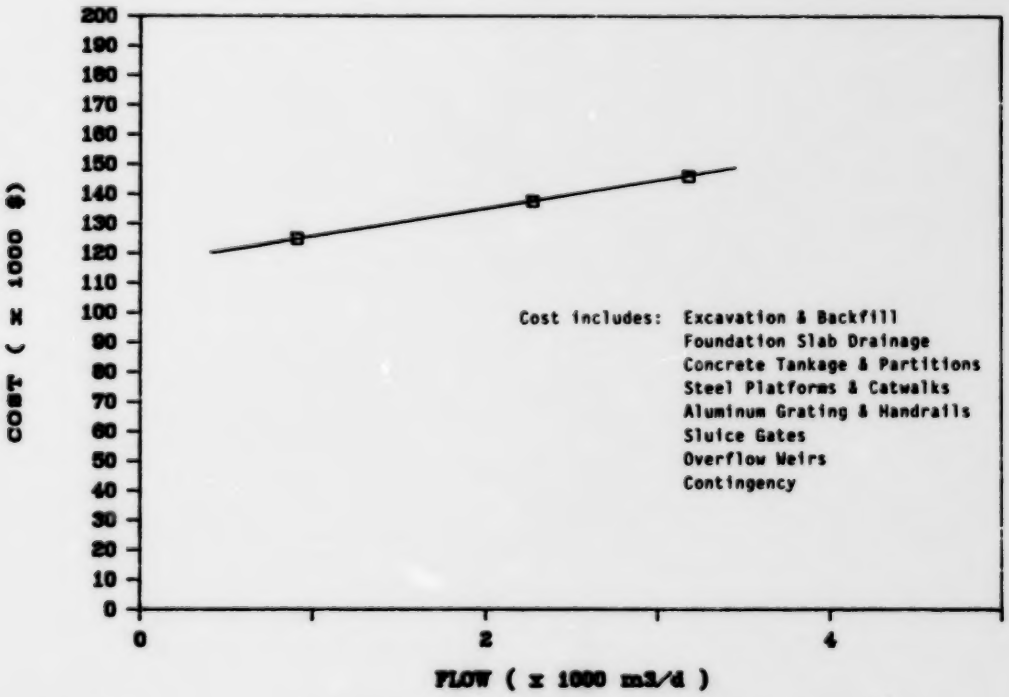

FIGURE A-6 EA BASIN RETROFIT TO A/O COST - CAPITAL

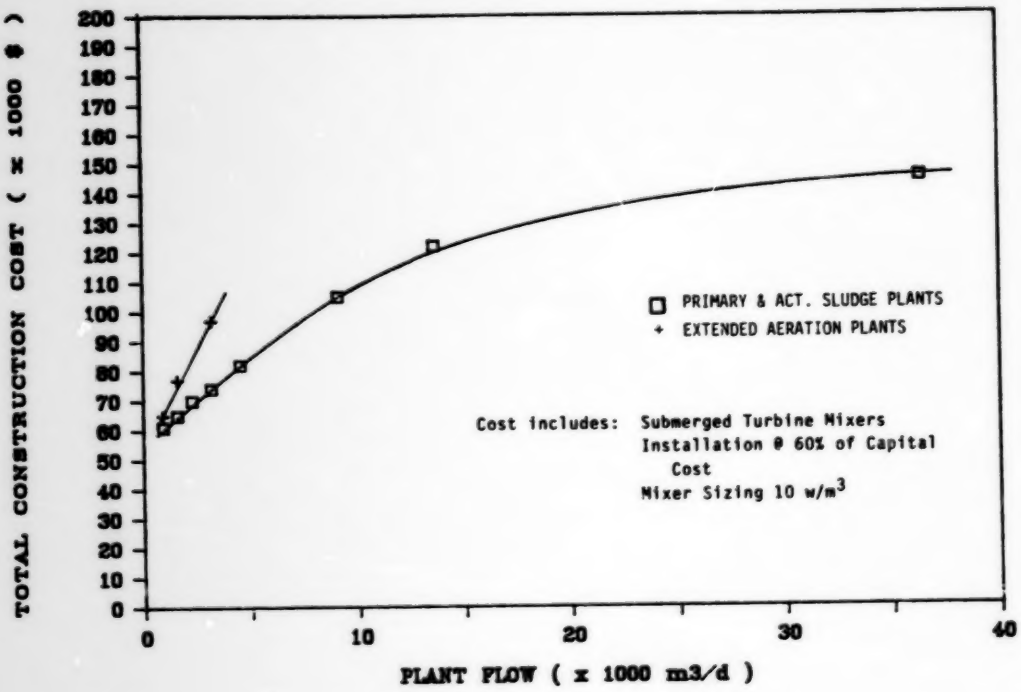

FIGURE A-7 A/O PROCESS MIXER COST - CAPITAL

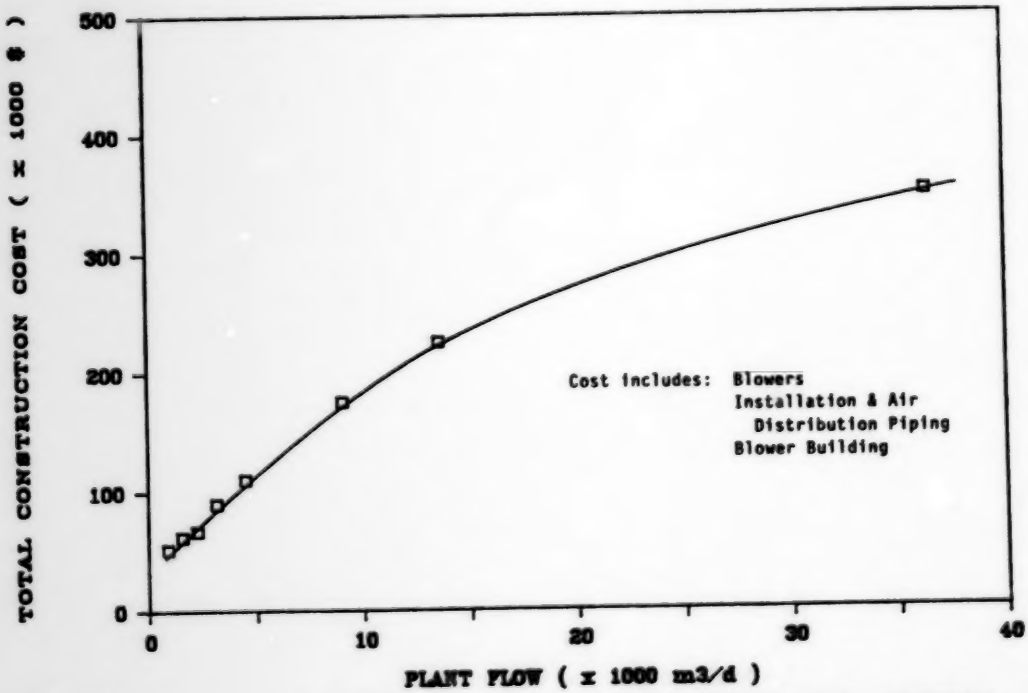

FIGURE A-8 A/O PROCESS BLOWER FACILITIES COST - CAPITAL

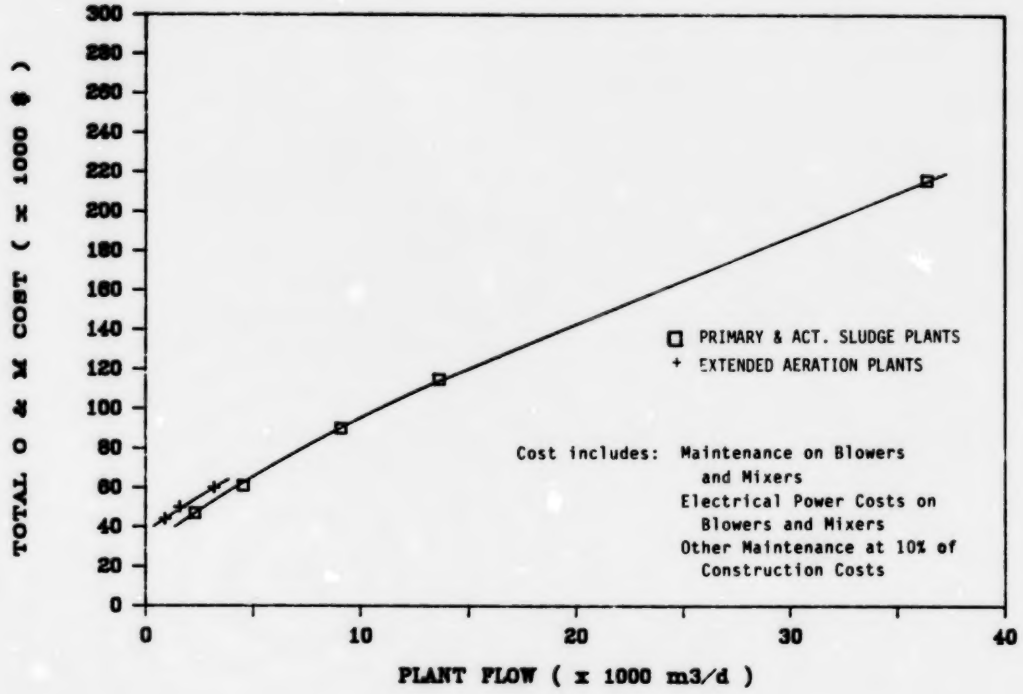

FIGURE A-9 A/O BASIN FACILITIES COST - O&M

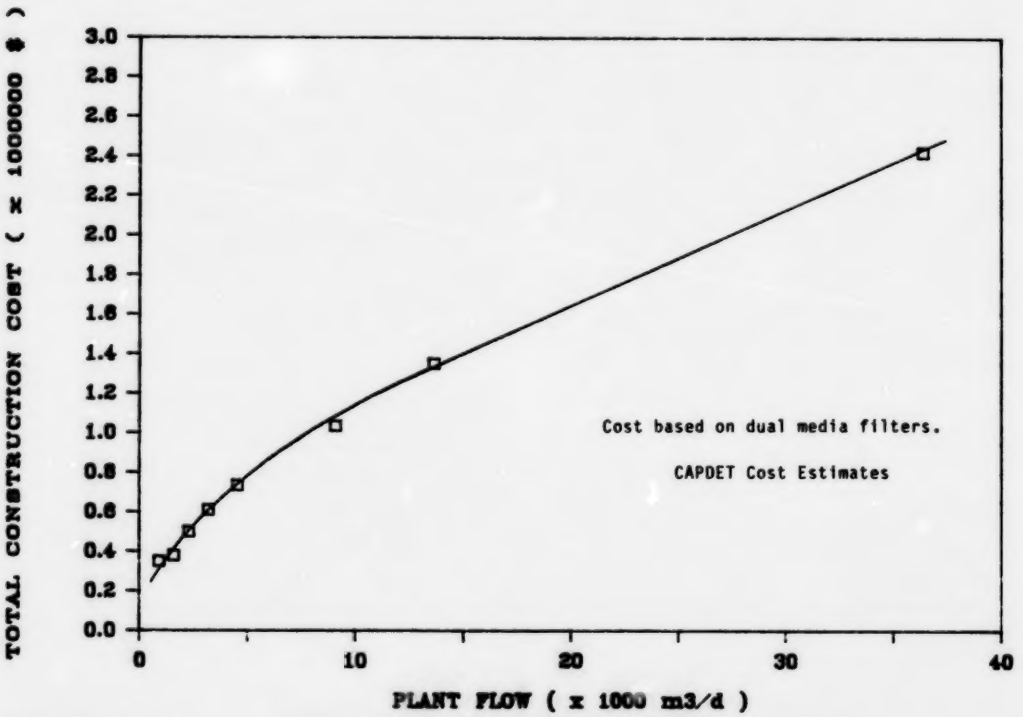

FIGURE A-10 FINAL EFFLUENT FILTER COST - CAPITAL

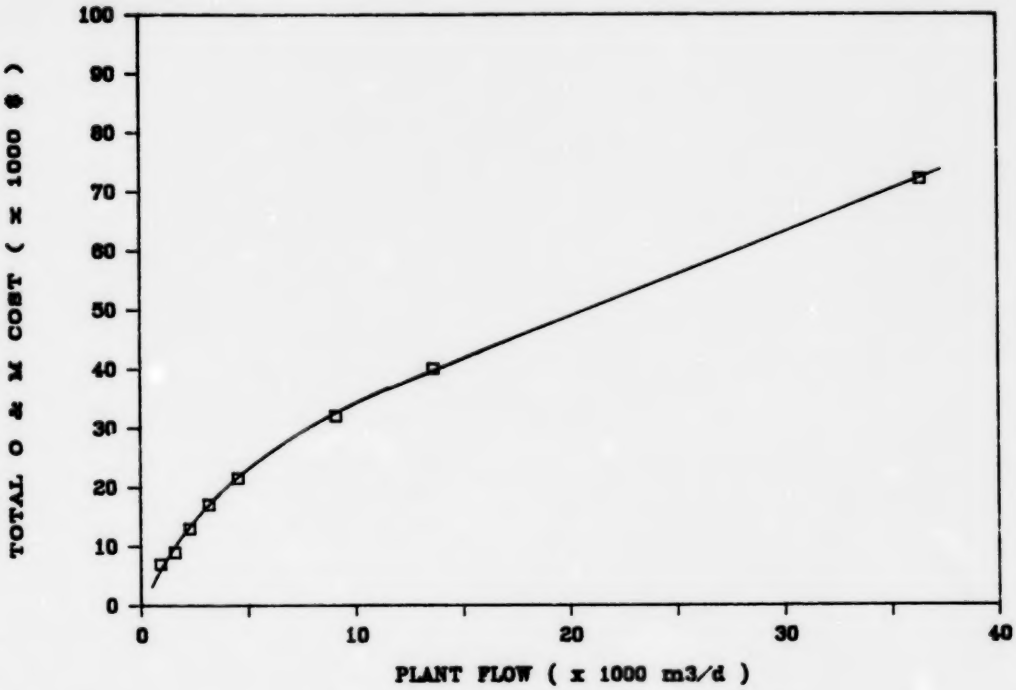

FIGURE A-11 FINAL EFFLUENT FILTER COST - O&M

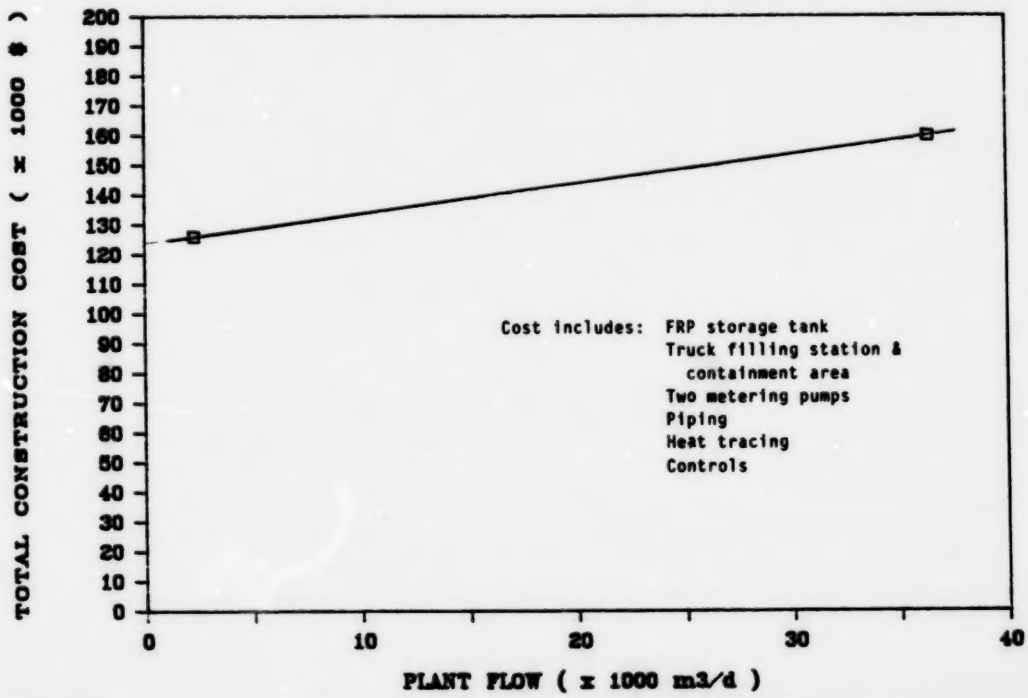

FIGURE A-12 CHEMICAL PHOSPHORUS REMOVAL COST - CAPITAL

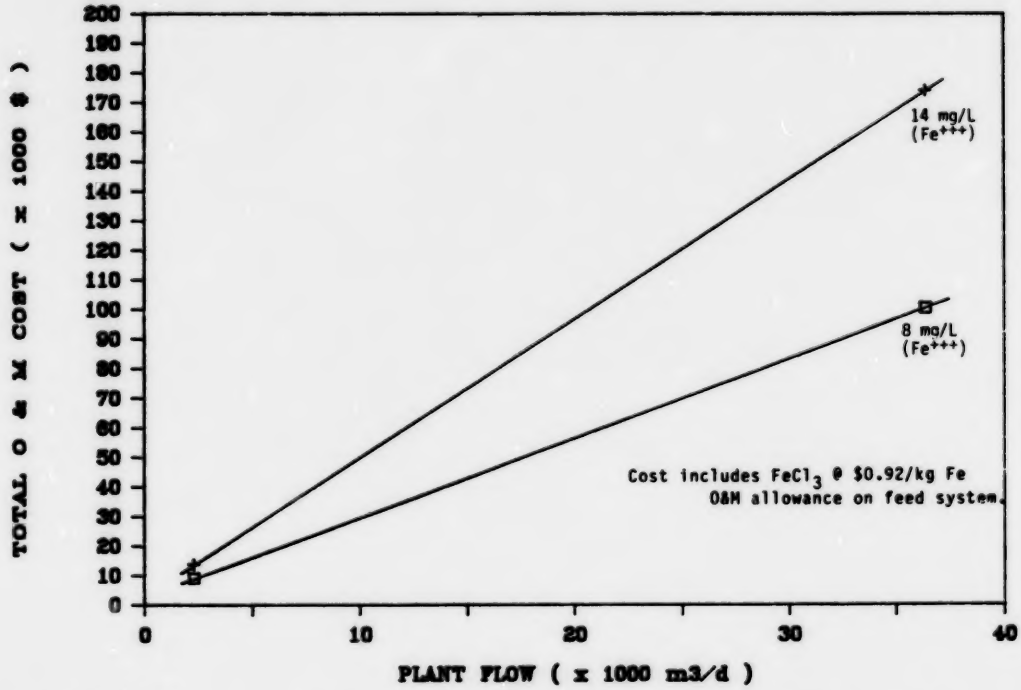

FIGURE A-13 CHEMICAL PHOSPHORUS REMOVAL COST - O&M

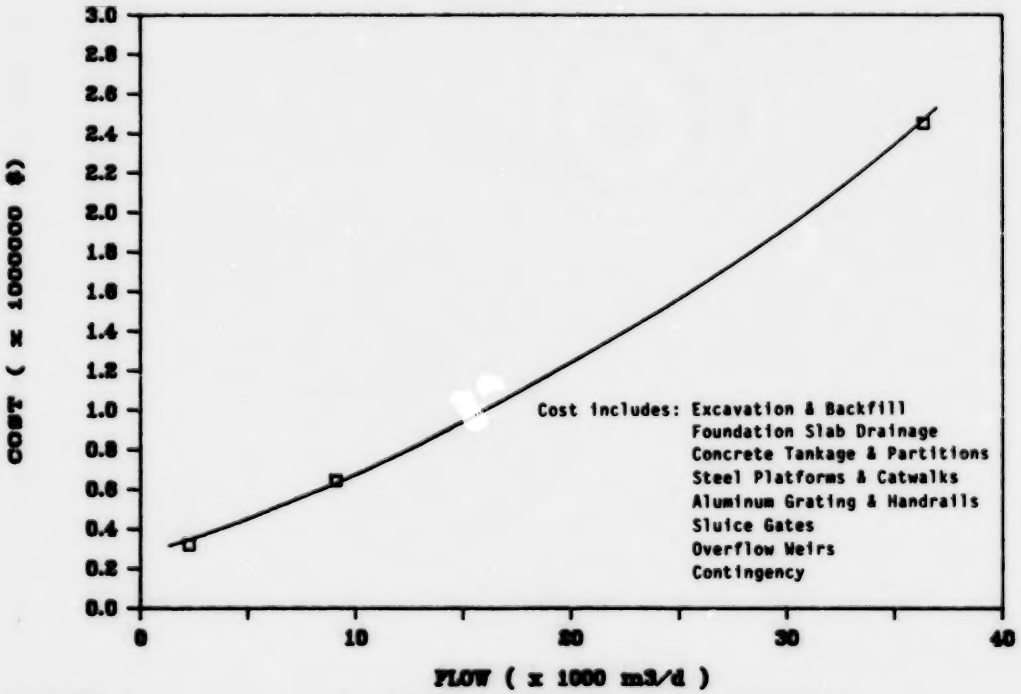

FIGURE A-14 NEW BARDENPHO BASIN COST - CAPITAL

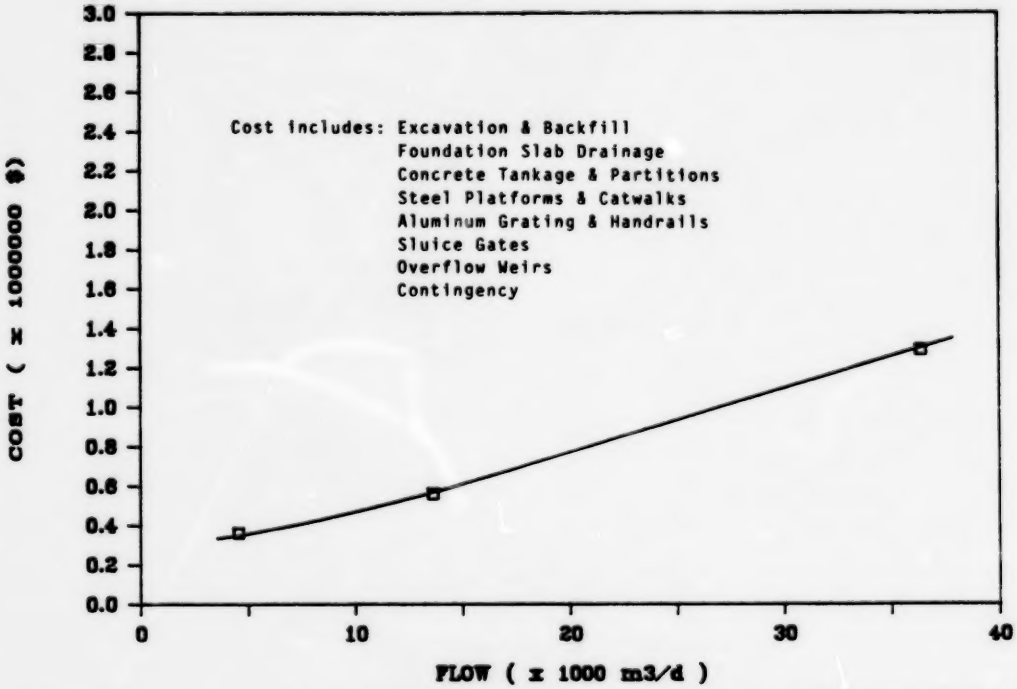

FIGURE A-15 CAS AERATION BASIN RETROFIT TO BARDENPHO PROCESS COST - CAPITAL

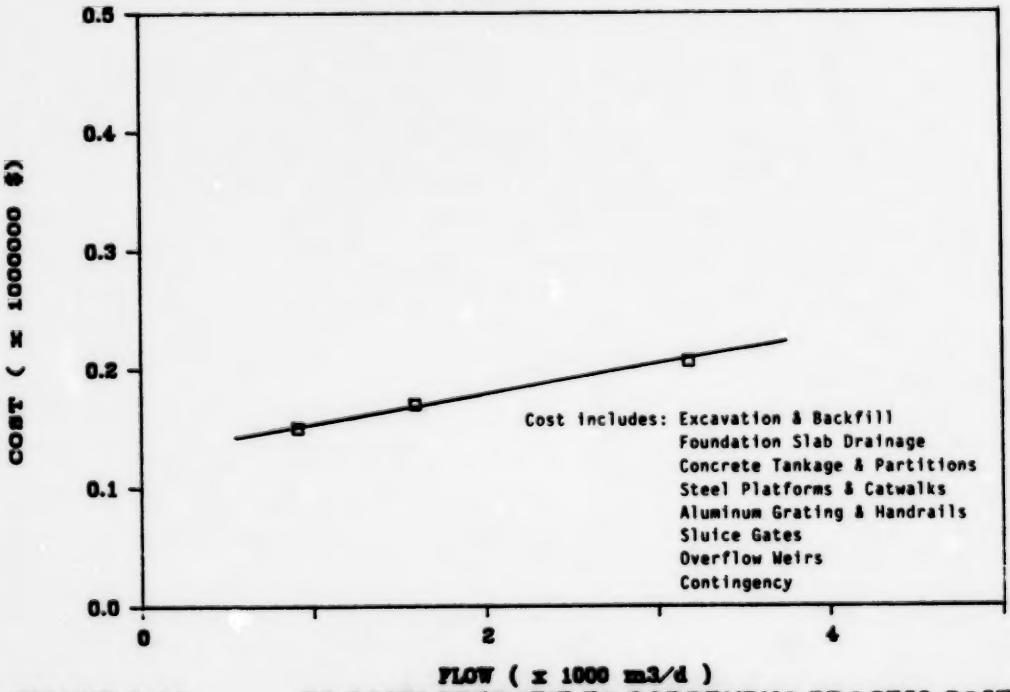

FIGURE A-16 EA BASIN RETROFIT TO BARDENPHO PROCESS COST - CAPITAL

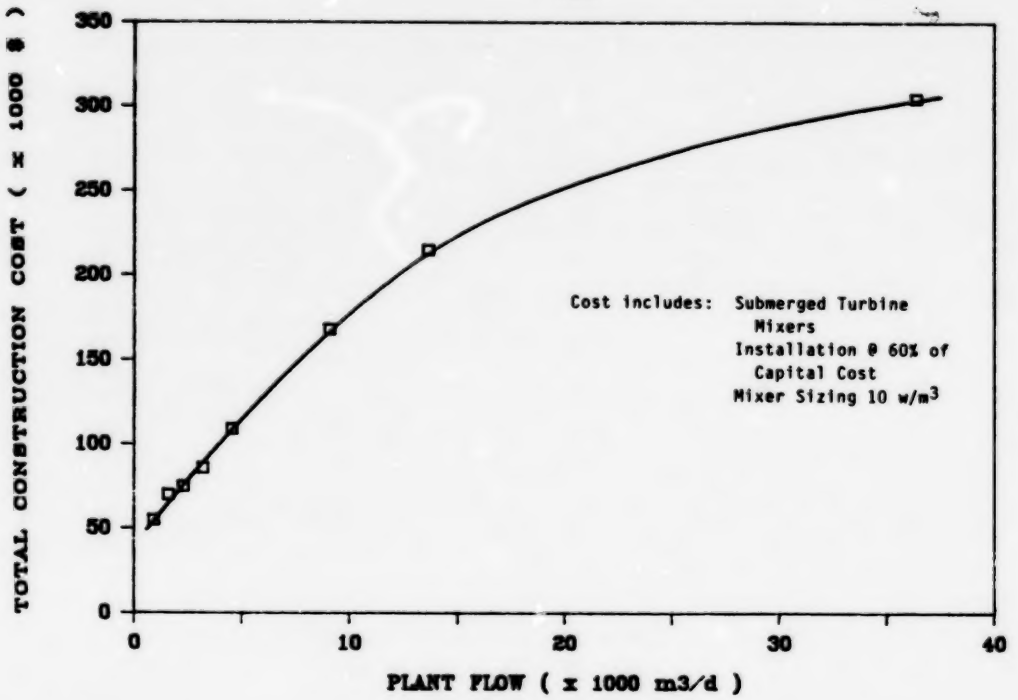

FIGURE A-17 BARDENPHO PROCESS MIXER COST - CAPITAL

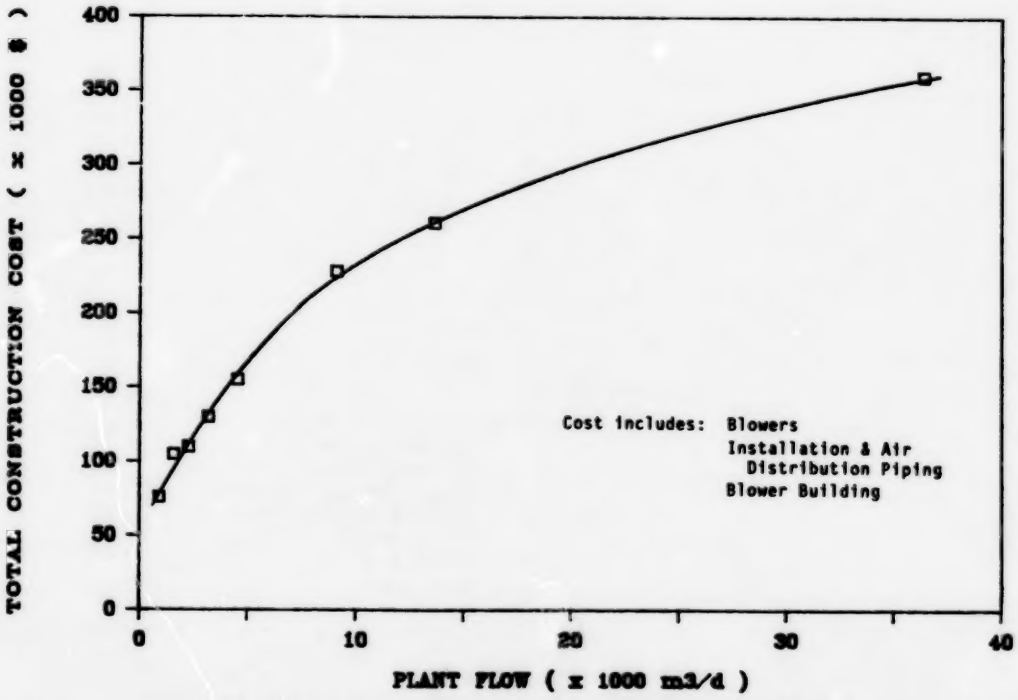

FIGURE A-18 BARDENPHO/UCT PROCESS BLOWER FACILITIES COST - CAPITAL

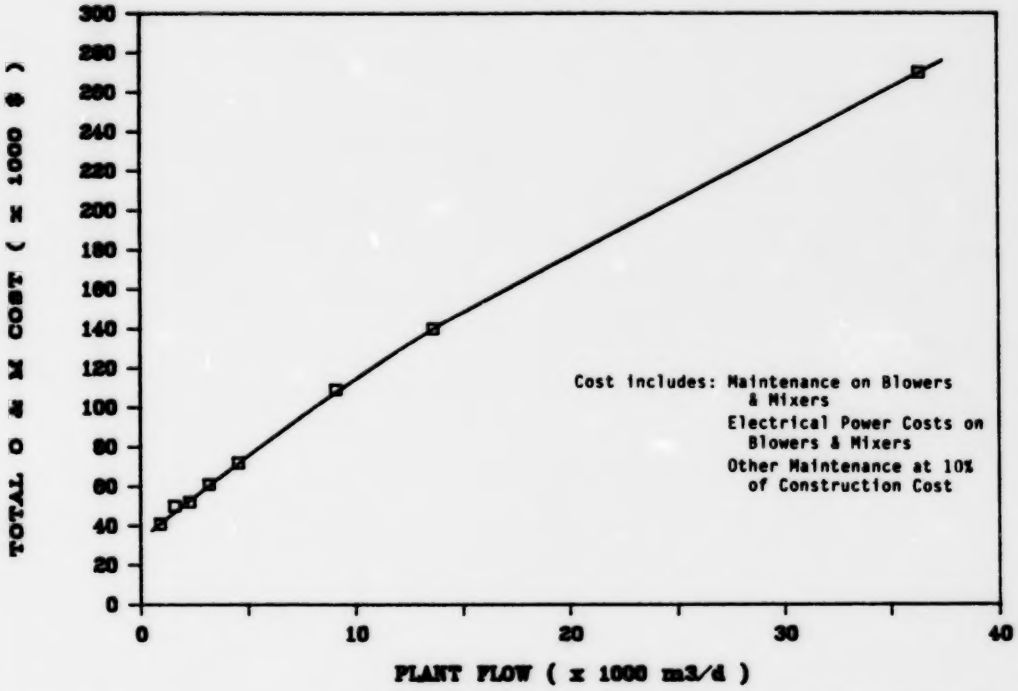

FIGURE A-19 BARDENPHO/UCT PROCESS BLOWER FACILITIES COST - O&M

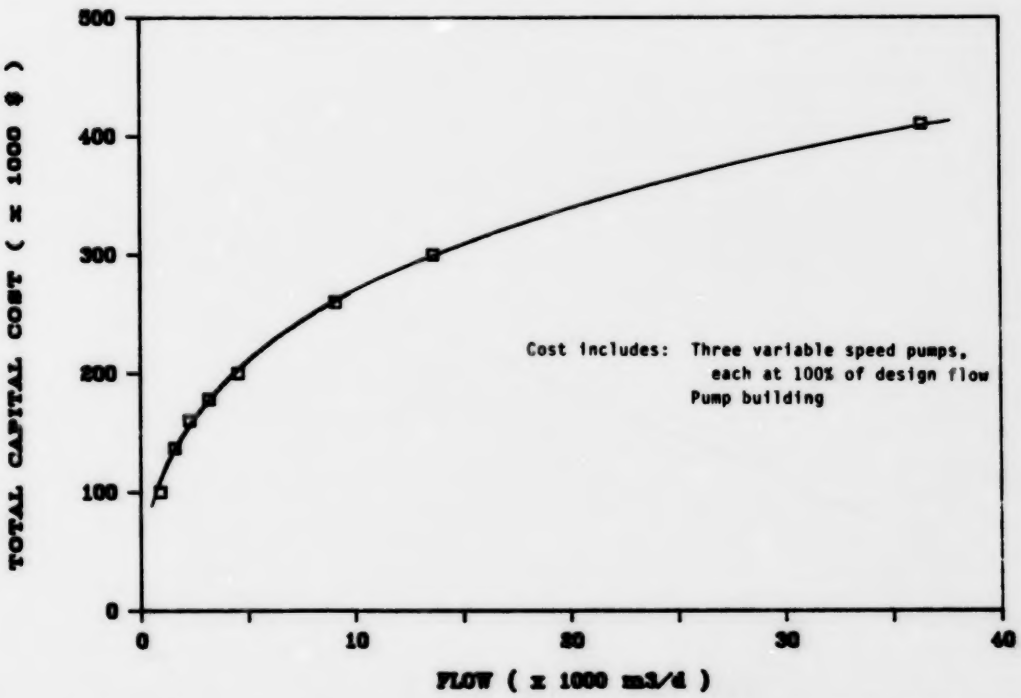

FIGURE A-20 BARDENPHO PROCESS INTERNAL RECYCLE COST - CAPITAL

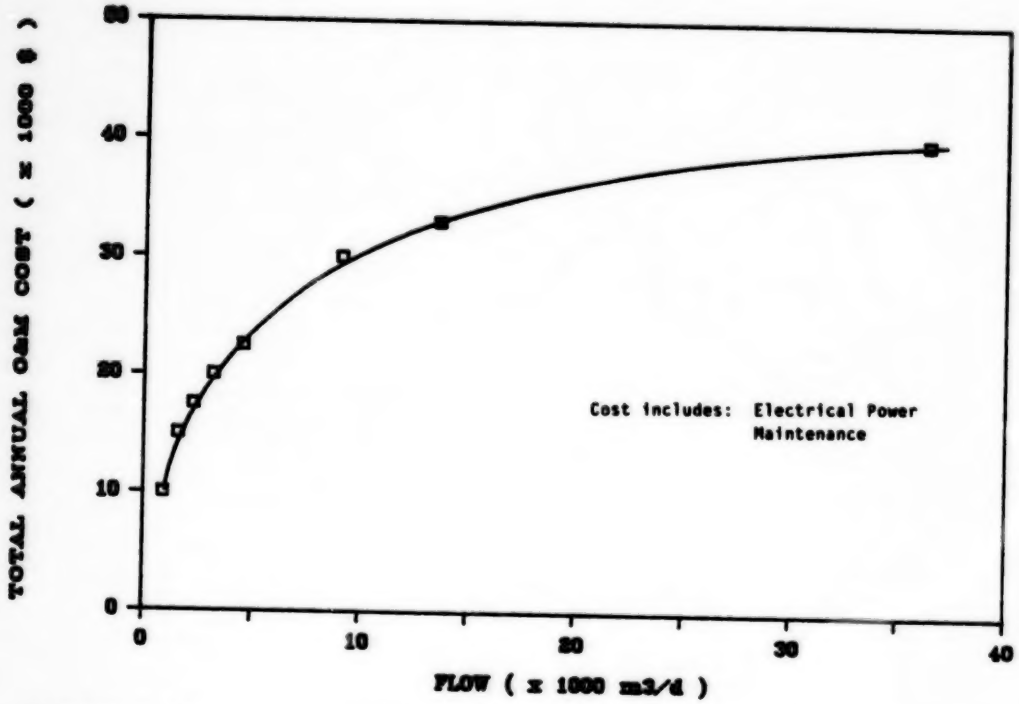

FIGURE A-21 BARDENPHO PROCESS INTERNAL RECYCLE COST - O&M

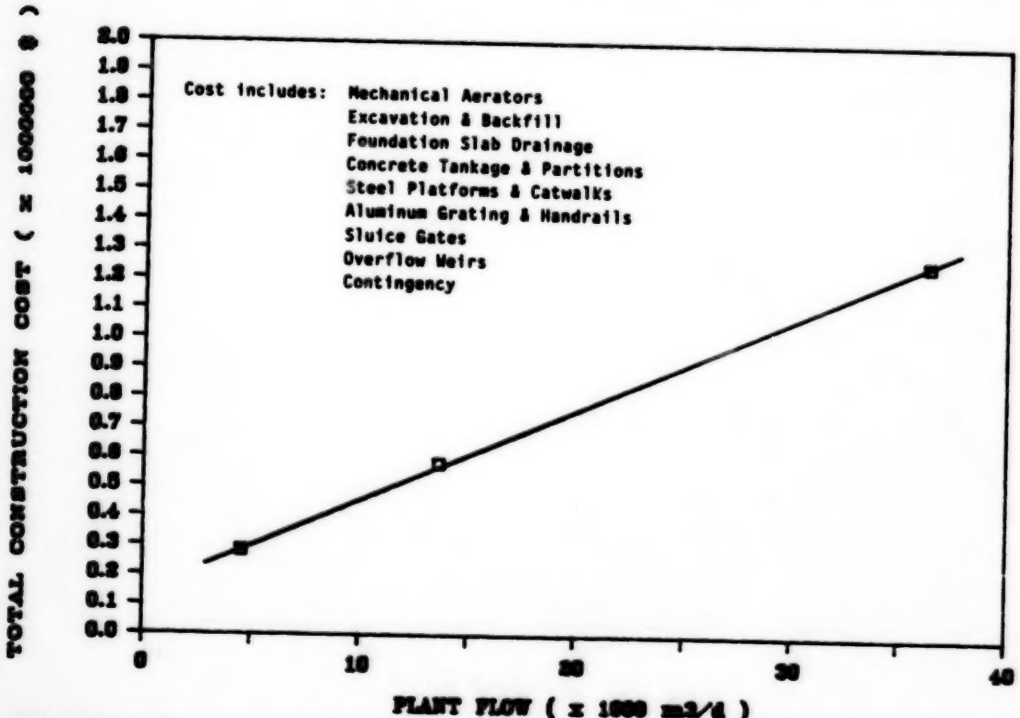

FIGURE A-22 CAS PROCESS AERATION BASIN COST - PROCESS

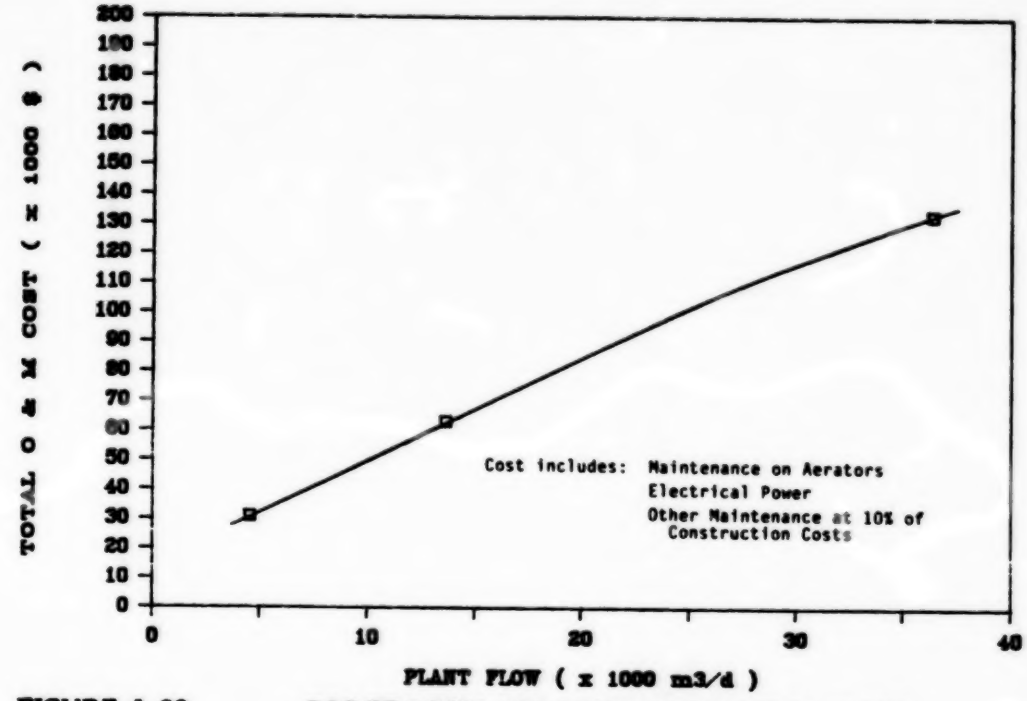

FIGURE A-23 CAS PROCESS AERATION BASIN COST - O&M

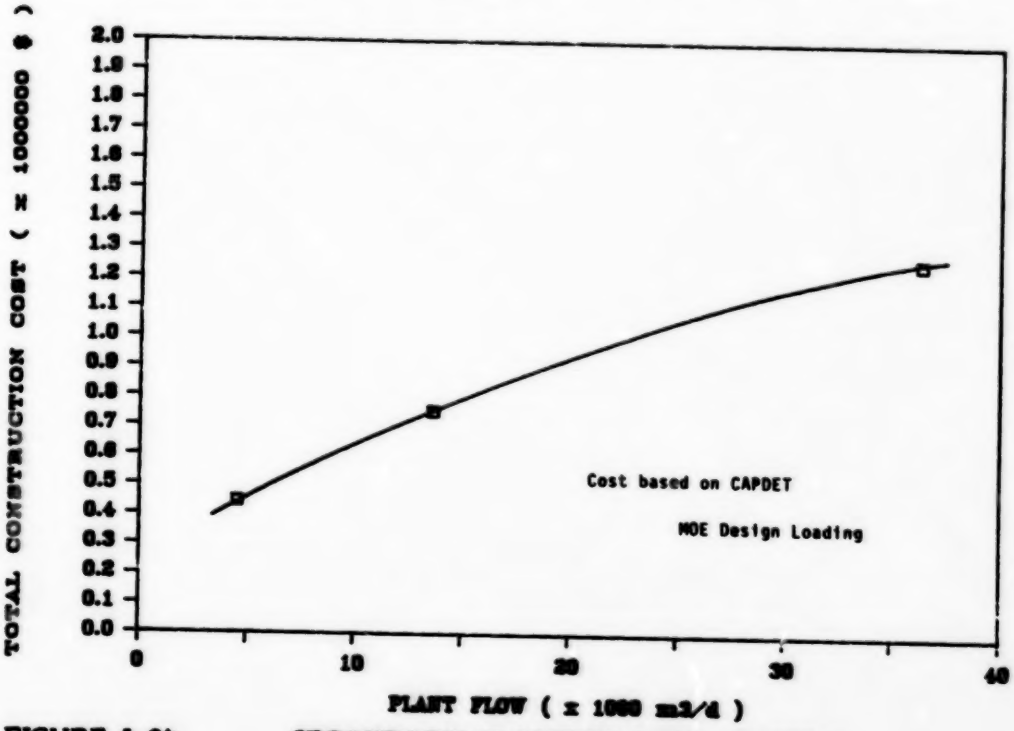

FIGURE A-24 SECONDARY CLARIFIER COST - CAPITAL

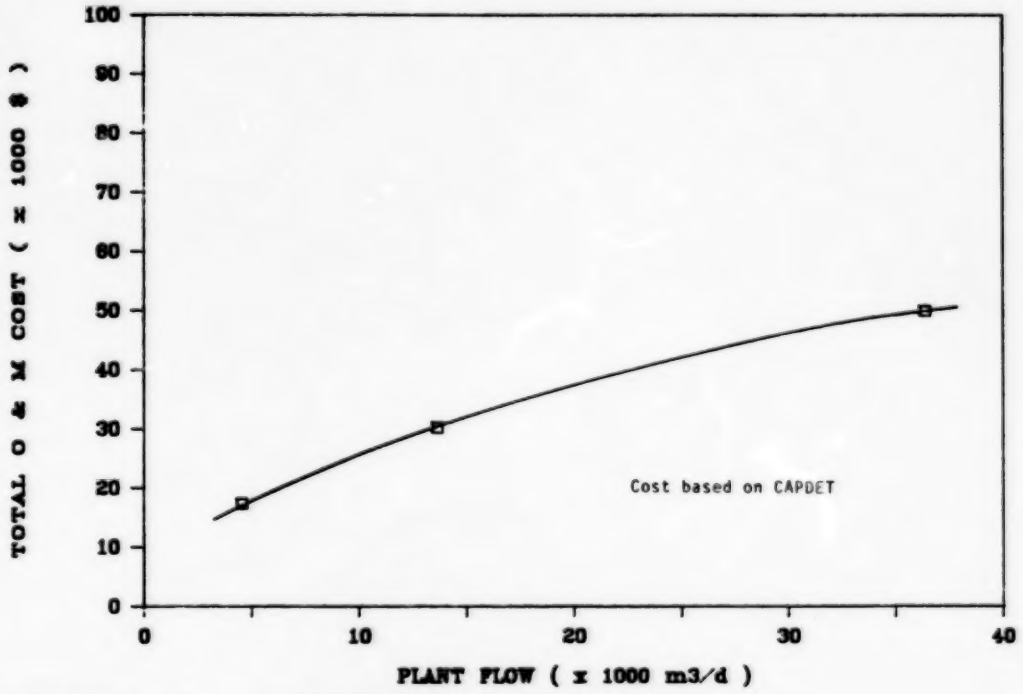

FIGURE A-25 SECONDARY CLARIFIER COST - O&M

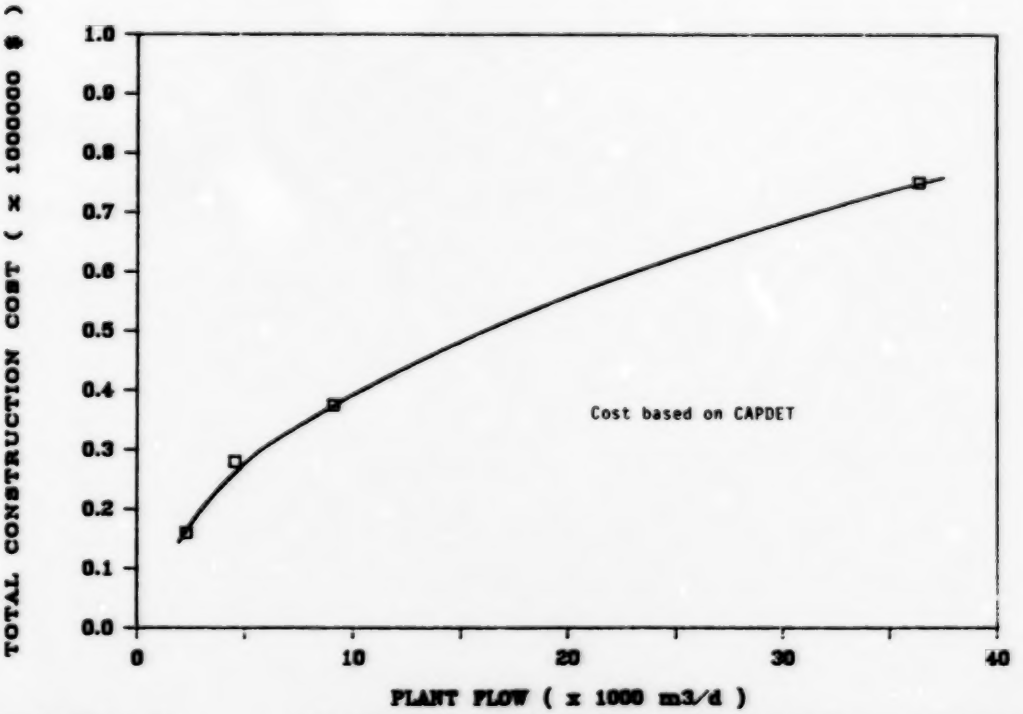

FIGURE A-26 PHOSTRIP PROCESS ANAEROBIC PHOSPHORUS STRIPPER TANK COST - CAPITAL

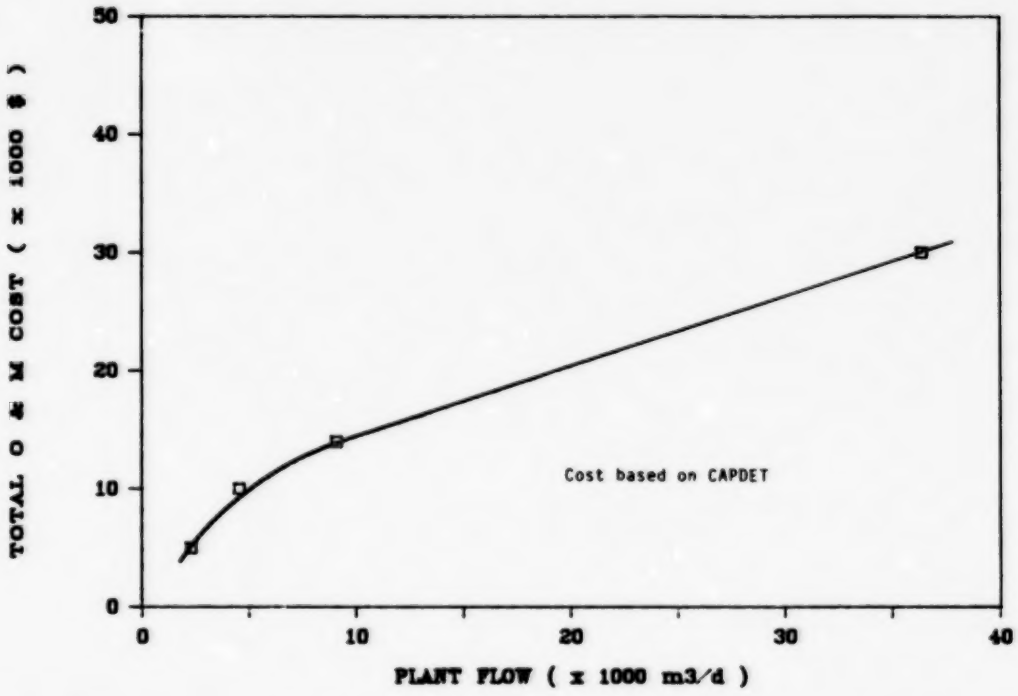

FIGURE A-27 PHOSTRIP PROCESS ANAEROBIC PHOSPHORUS STRIPPER TANK COST - O&M

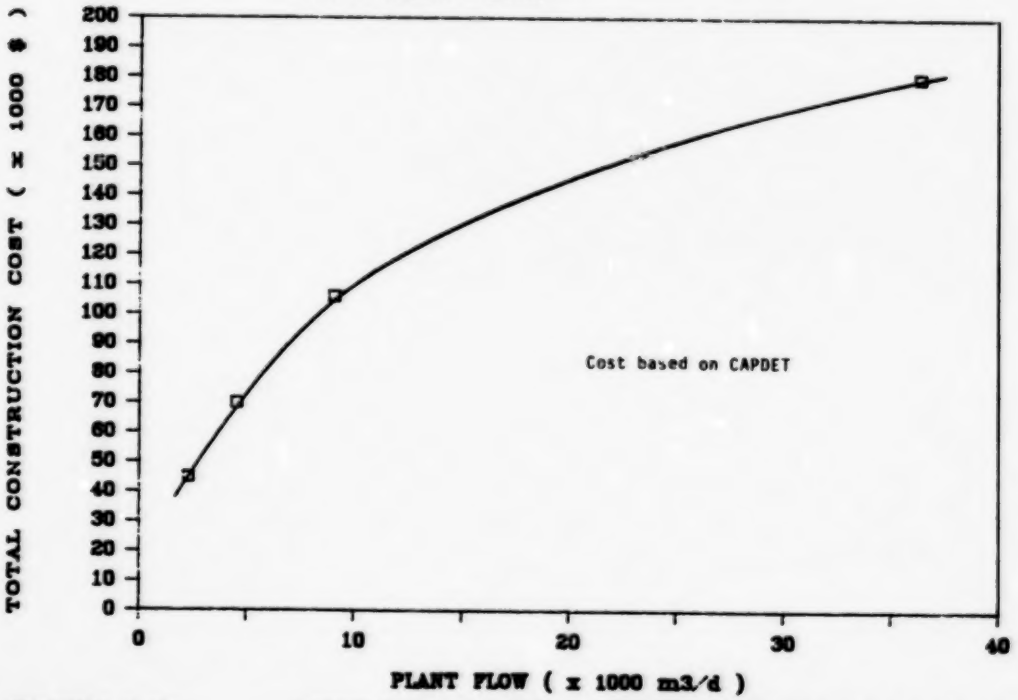

FIGURE A-28 PHOSTRIP PROCESS UNDERFLOW PUMPING COST - CAPITAL

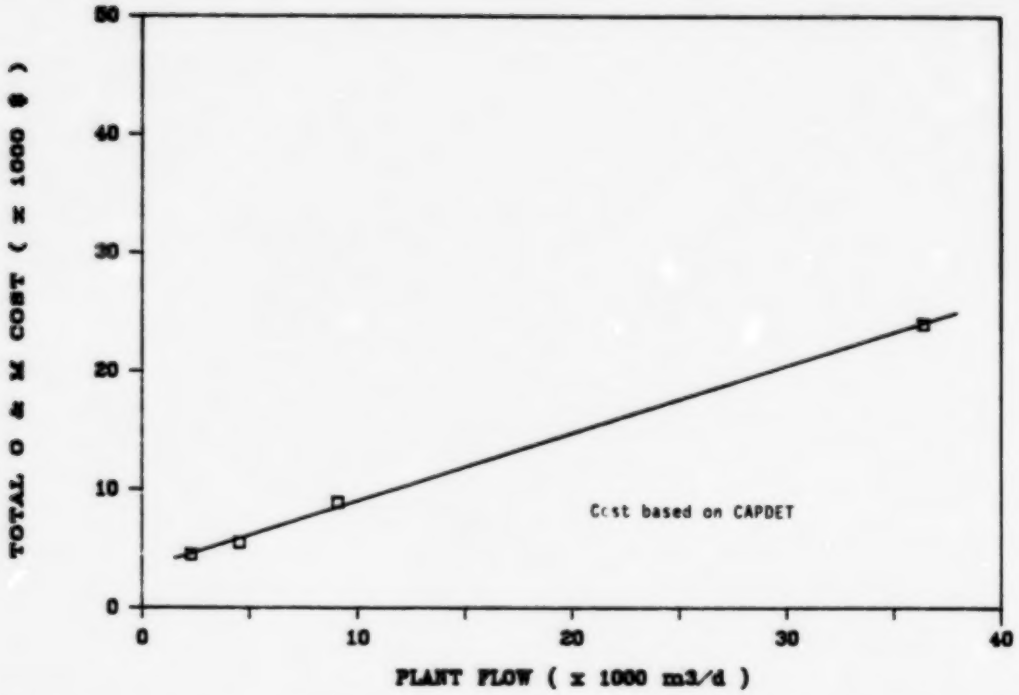

FIGURE A-29 PHOSTRIP PROCESS UNDERFLOW PUMPING COST - O&M

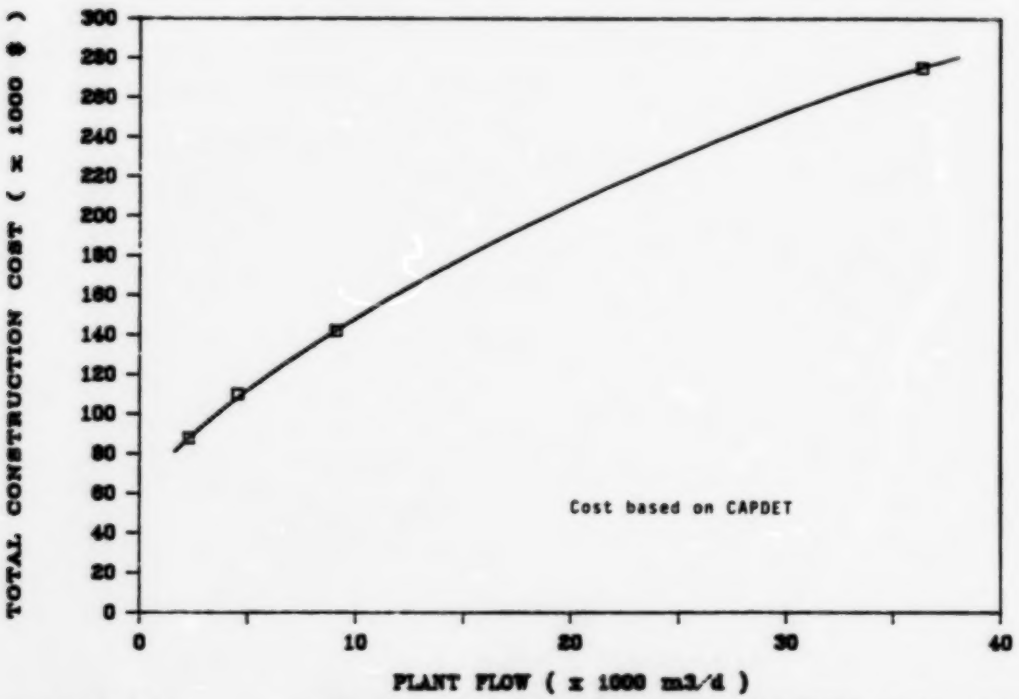

FIGURE A-30 PHOSTRIP PROCESS REACTOR CLARIFIER COST - CAPITAL

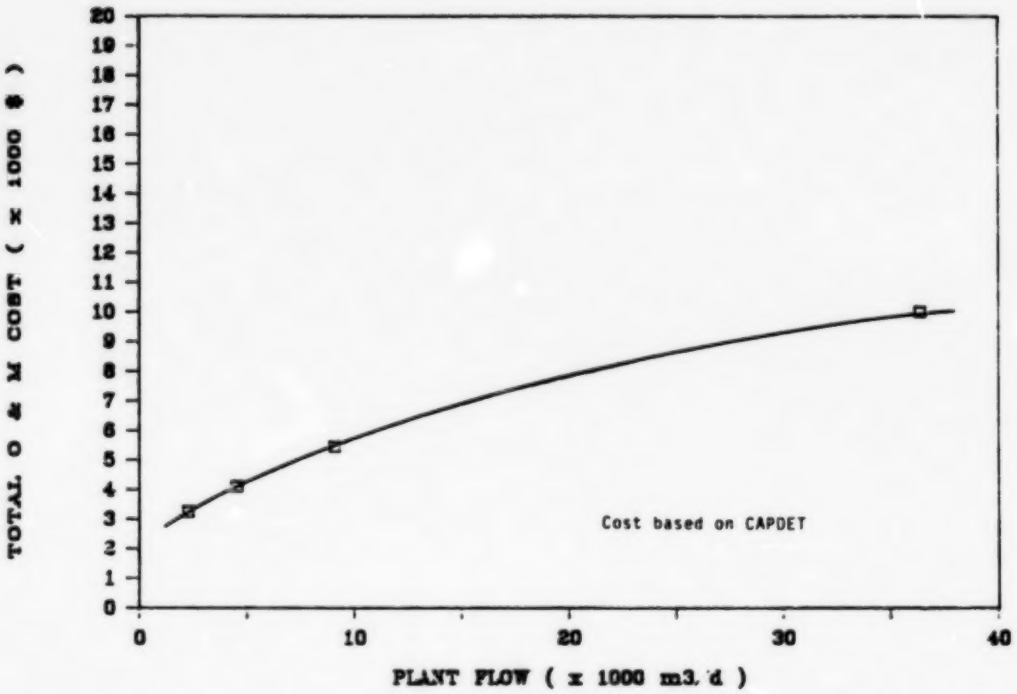

FIGURE A-31 PHOSTRIP PROCESS REACTOR CLARIFIER COST - O&M

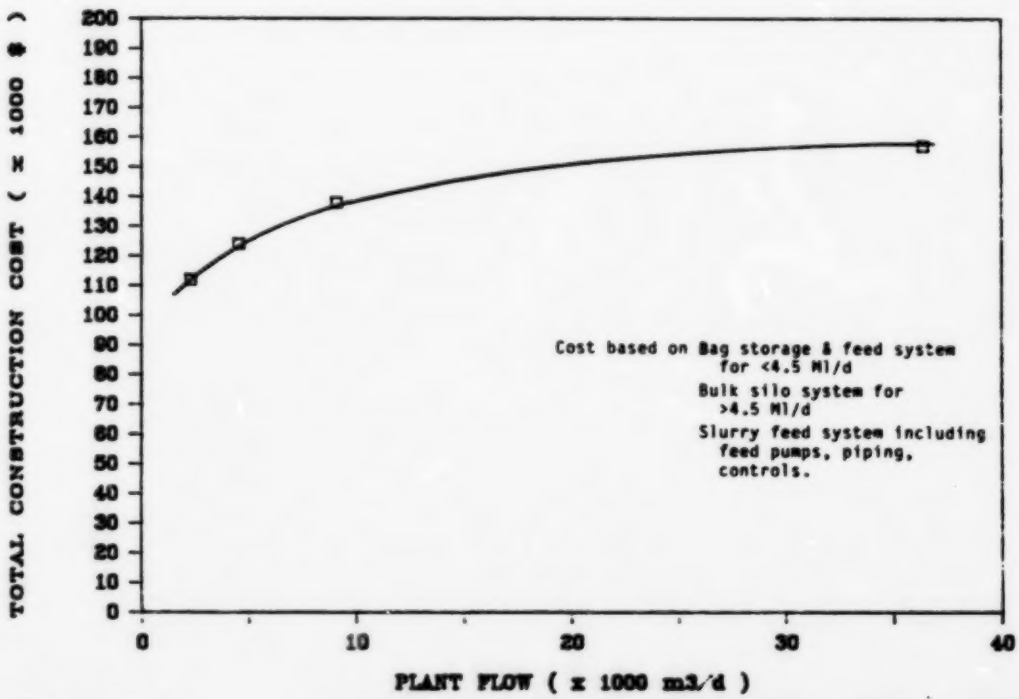

FIGURE A-32 PHOSTRIP PROCESS LIME FEED SYSTEM COST - CAPITAL

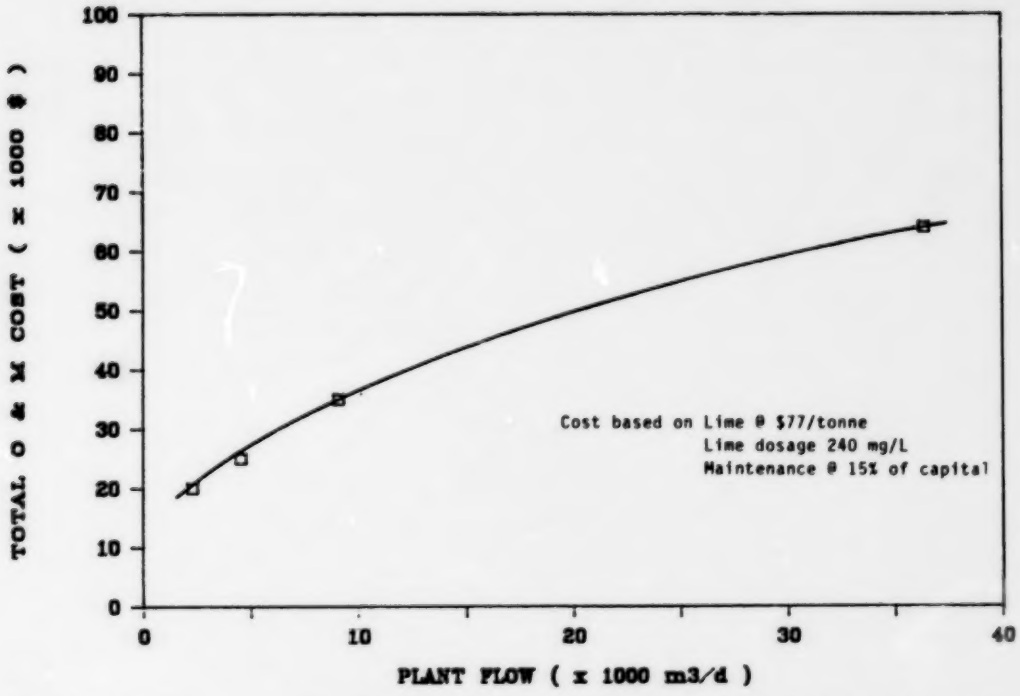

FIGURE A-33 PHOSTRIP PROCESS LIME FEED SYSTEM COST - O&M

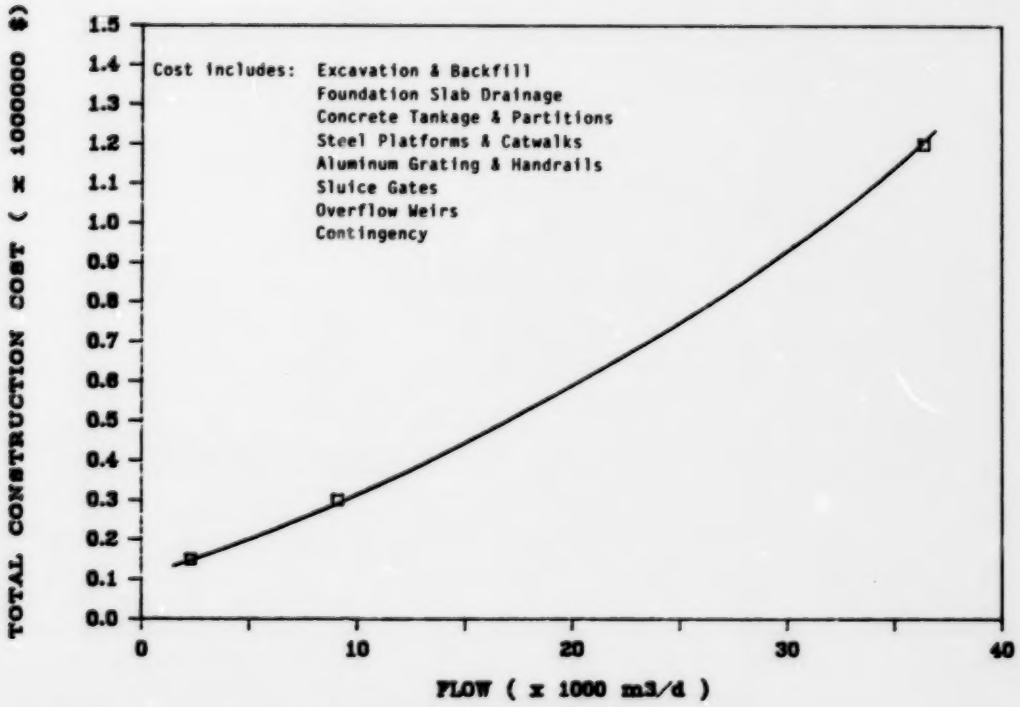

FIGURE A-34 NEW UCT BASIN COST - CAPITAL

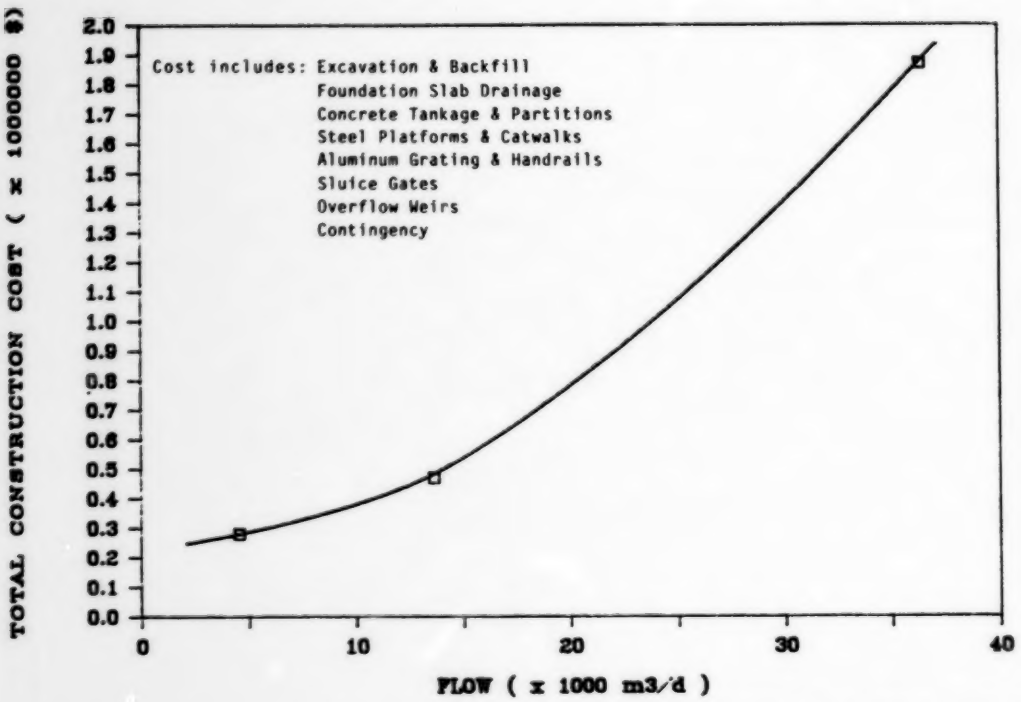

FIGURE A-35 CAS AERATION RETROFIT TO UCT PROCESS COST - CAPITAL

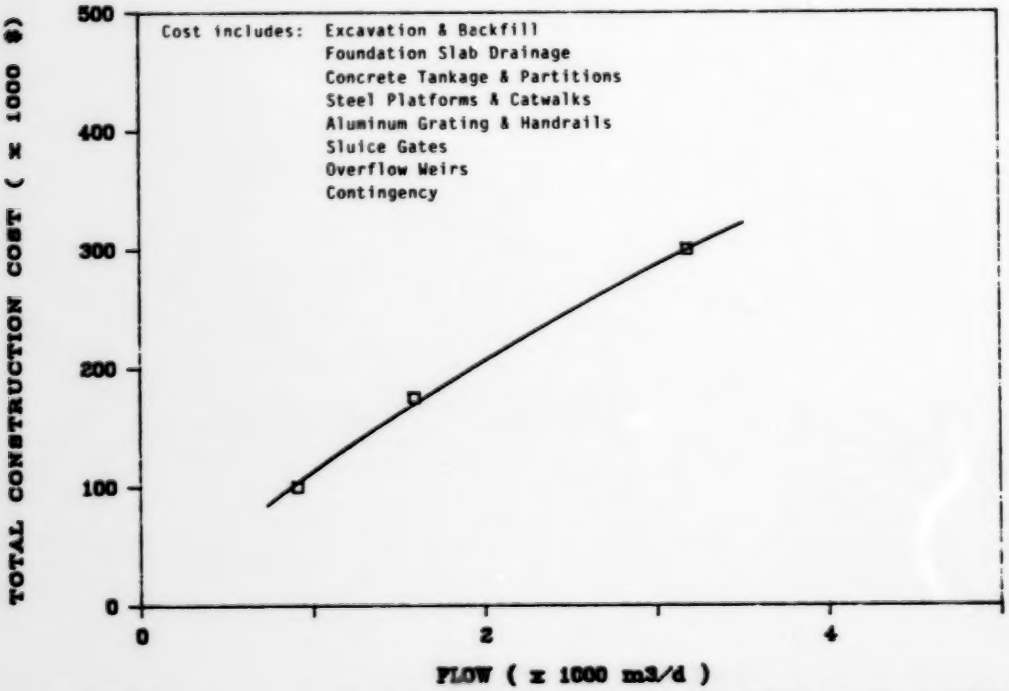

FIGURE A-36 EA BASIN RETROFIT TO UCT PROCESS COST - CAPITAL

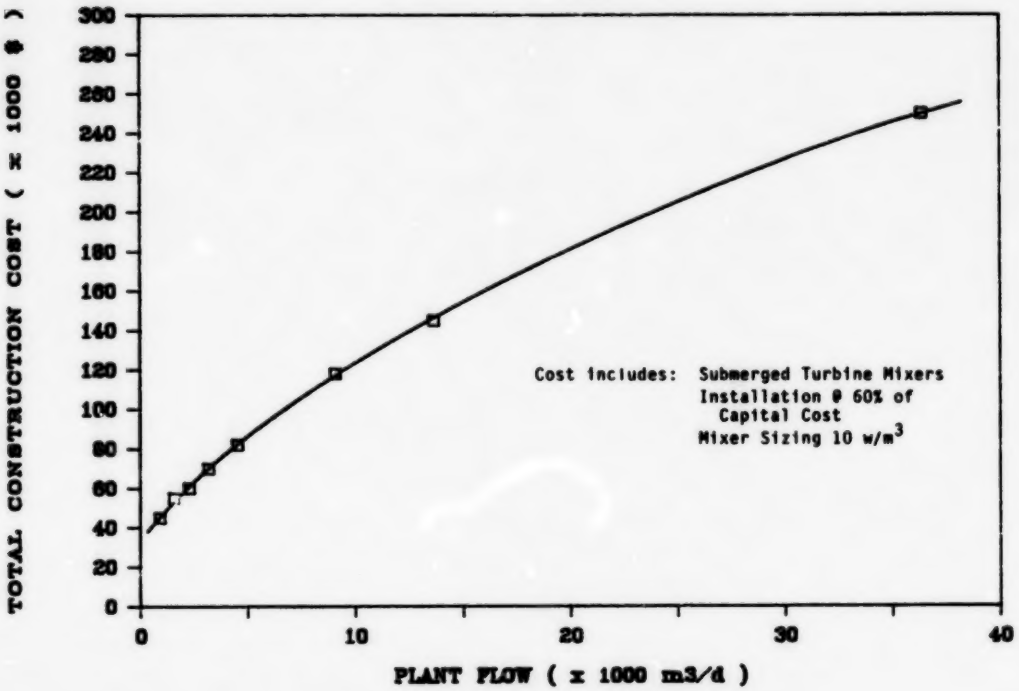

FIGURE A-37 UCT PROCESS MIXER COST - CAPITAL

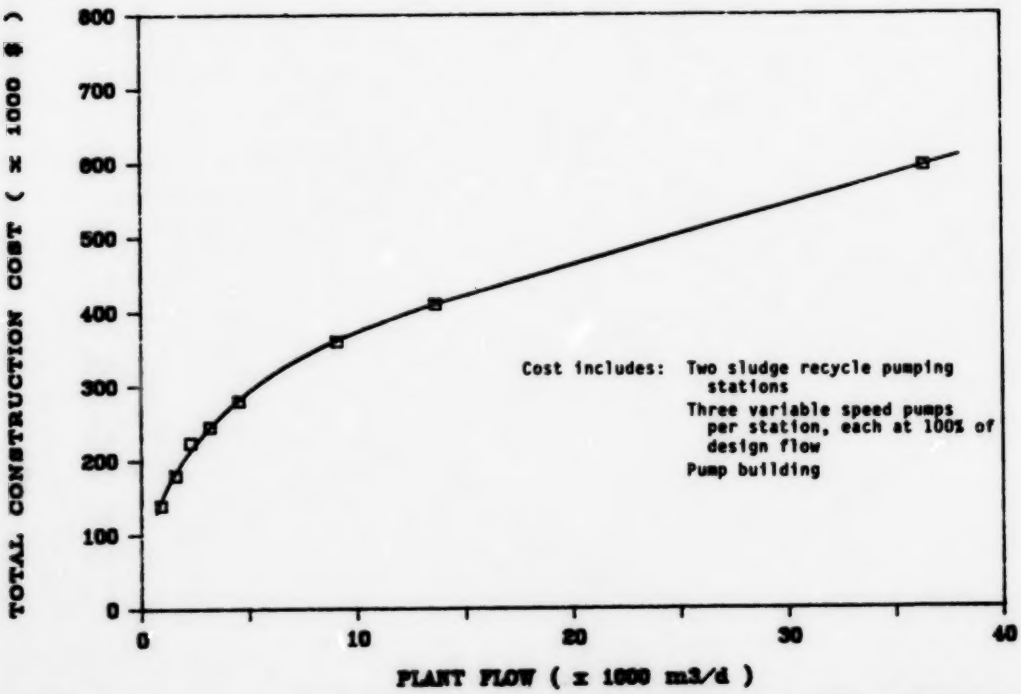

FIGURE A-38 UCT PROCESS INTERNAL RECYCLE COST - CAPITAL

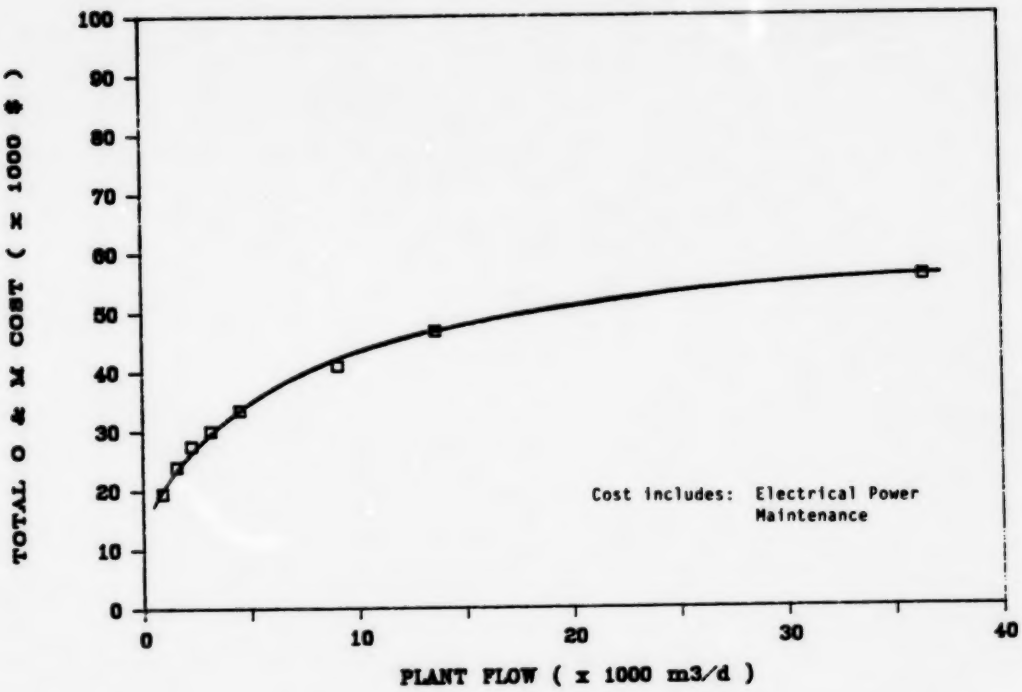

FIGURE A-39 UCT PROCESS INTERNAL RECYCLE COST - O&M

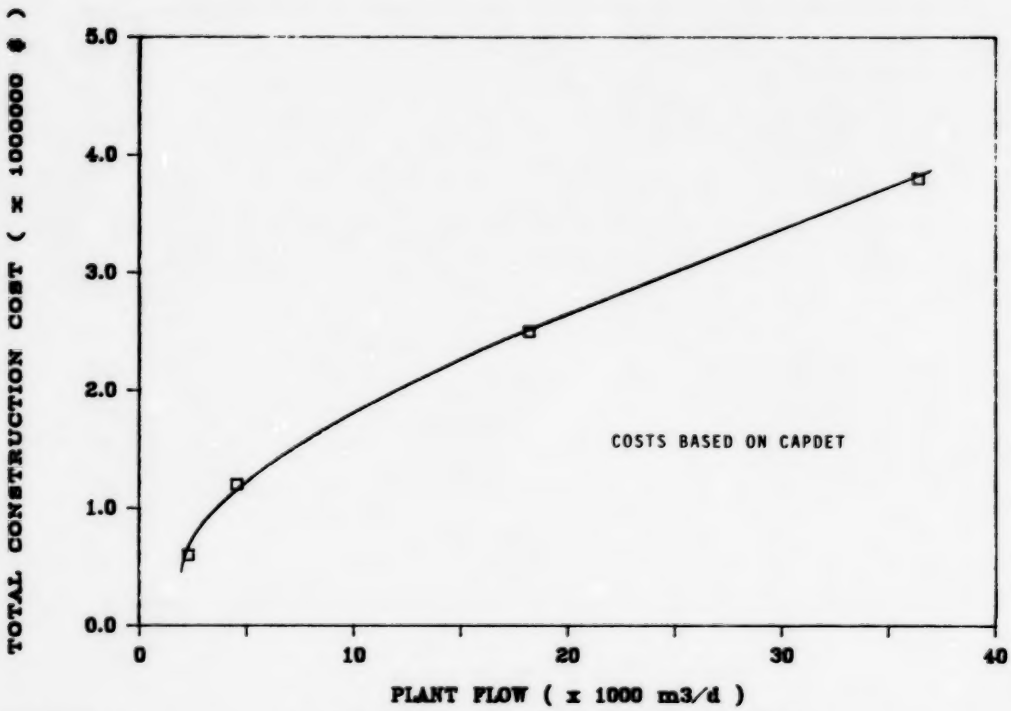

FIGURE B-1 FLOW EQUALIZATION COST - CAPITAL

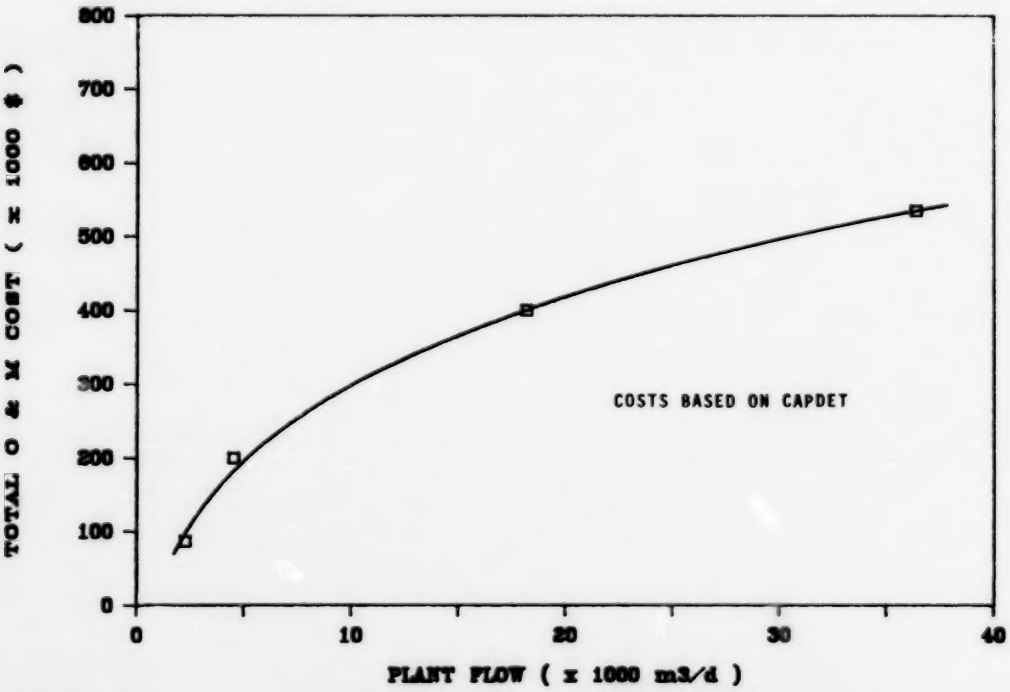

FIGURE B-2 FLOW EQUALIZATION COST - O&M

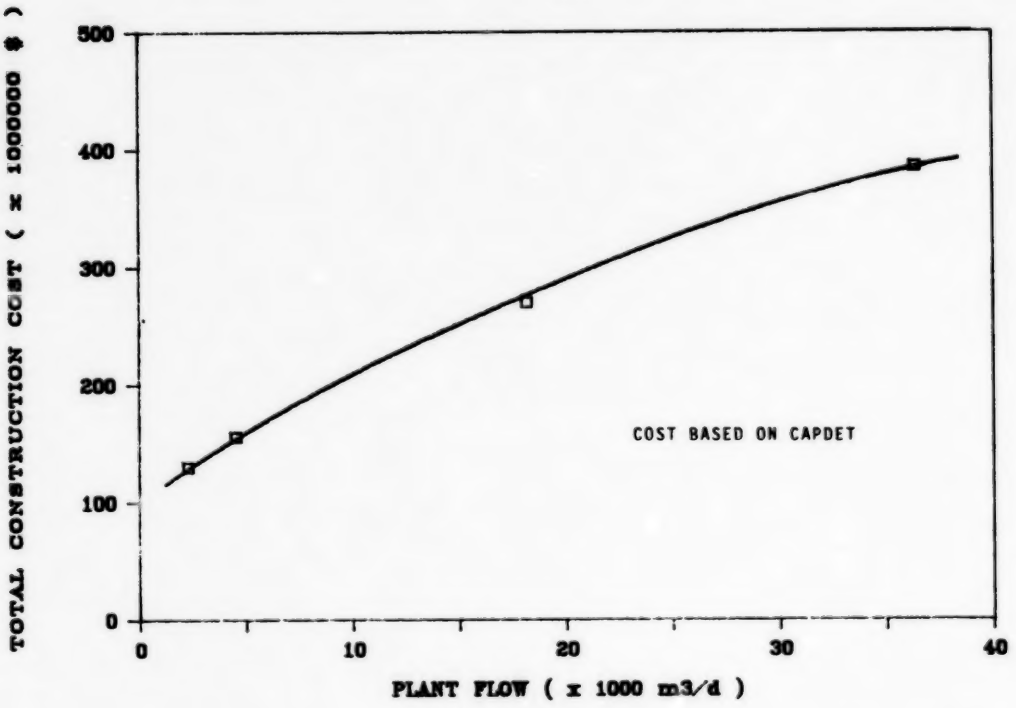

FIGURE B-3 CHLORINE CONTACT CHAMBER COST - CAPITAL

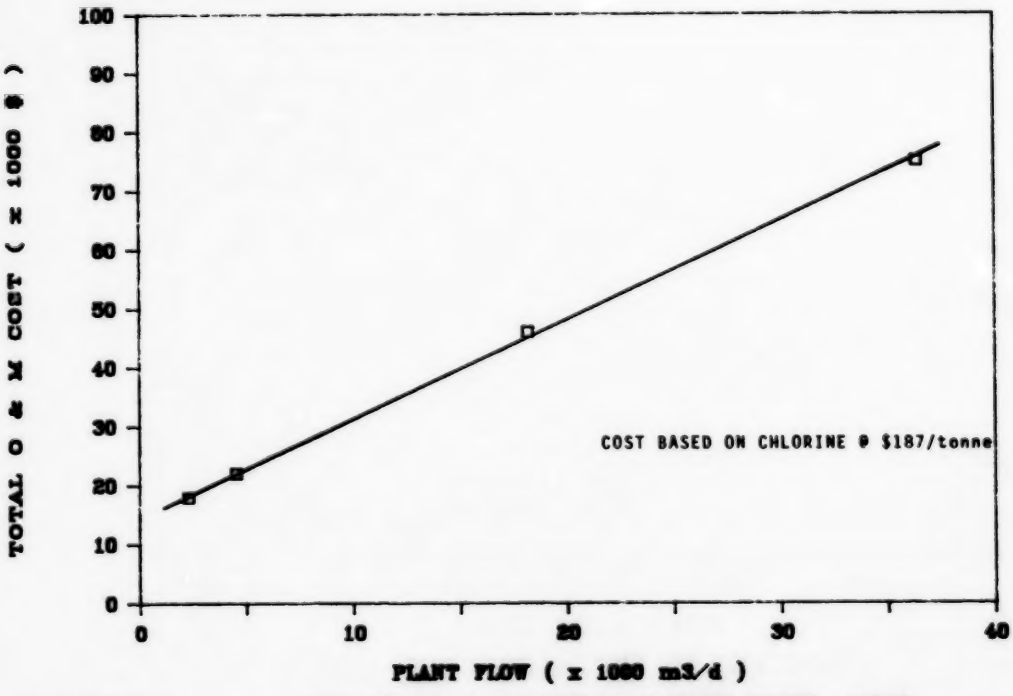

FIGURE B-4 CHLORINE CONTACT CHAMBER COST - O&M

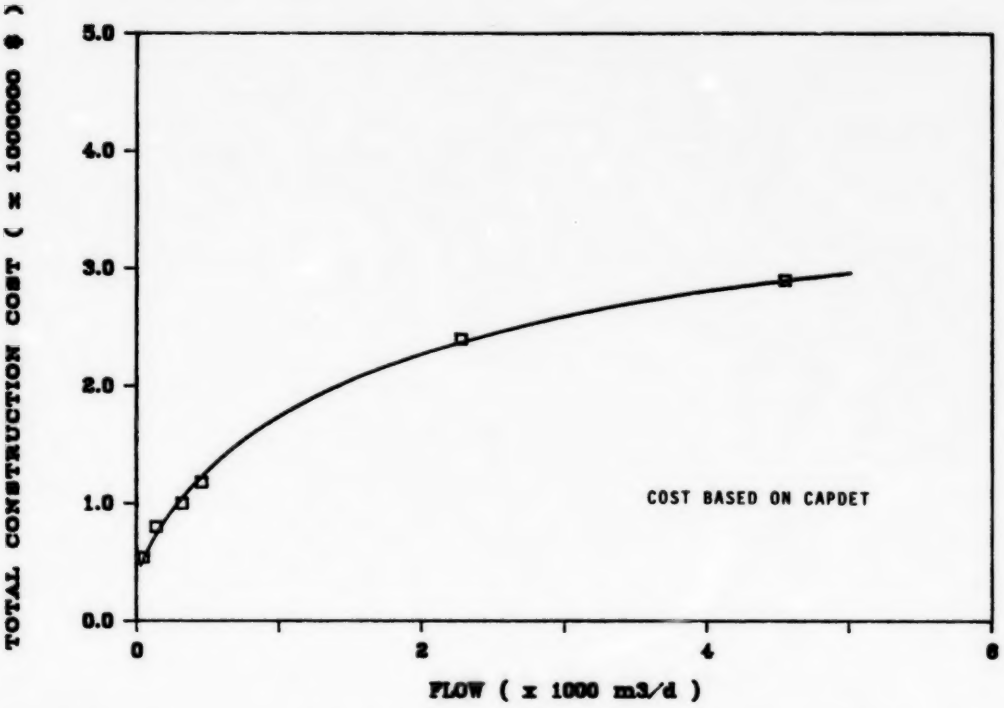

FIGURE B-5 DISSOLVED AIR FLOTATION COST - CAPITAL

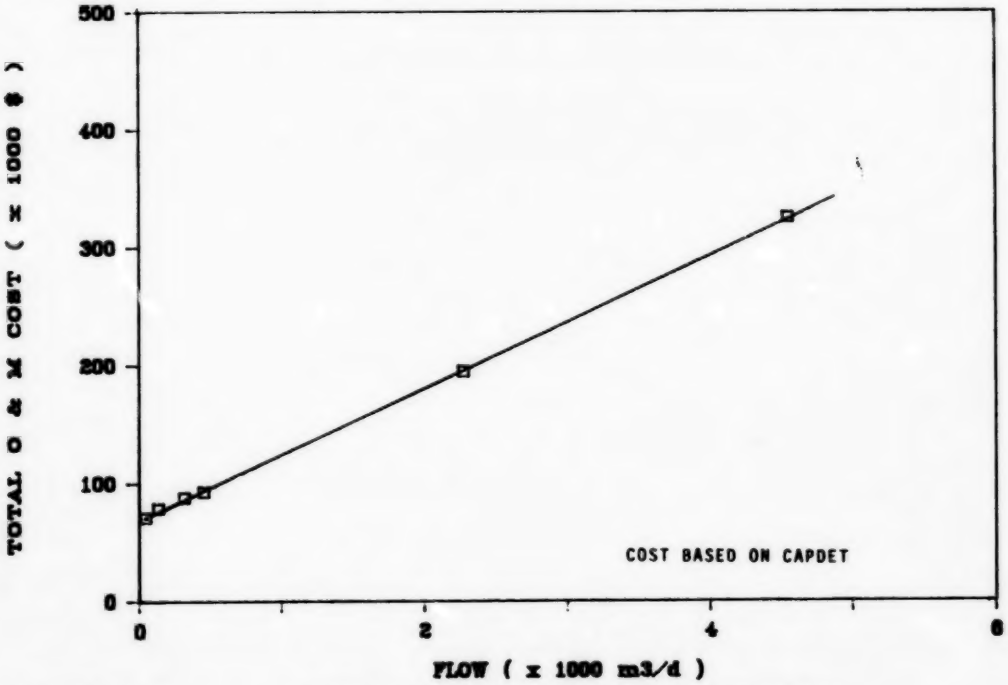

FIGURE B-6 DISSOLVED AIR FLOTATION COST - O&M